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# Current Status of Commercial Utility Flue Gas Desulfurization Systems

Radian Corp., Austin, TX

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# THE CURRENT STATUS OF COMMERCIAL UTILITY FLUE GAS DESULFURIZATION SYSTEMS

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The report discusses the current status of commercial flue gas desulfurization (FGD) processes applied to coal-fired utility boilers in the U.S. Major objectives of the work were to examine the impacts of the 1979 New Source Performance Standards on FGD system design and operation, and to identify recent improvements in the technology. In the 4 years since promulgation of the NSPS, the wet limestone process has been selected by utilities for 75% of the new plant capacity. During this time, 77 plants representing over 37,000 MW of capacity have selected FGD systems. Several major trends in the design of limestone systems have become fairly standardized. Nearly all new systems are being built with spare absorber modules, to qualify for the NSPS emergency bypass provisions. The predominant absorber design is the open spray tower, due to minimal maintenance requirements. Forced oxygen to produce gypsum solids, which can then be landfilled, is being incorporated in many new units. The spray drying process has been selected for 15% of the new sites in the last 4 years. The remaining throwaway and regenerable systems have not experienced any significant increases in application. Finally, organic acid addition has been successfully demonstrated on limestone systems to improve SO2 removal and system reliability. It is being used at two sites to upgrade older systems.

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| Coal | | 21D |
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# ABSTRACT

This report discusses the current status of commercial flue gas desulfurization processes applied to coal-fired utility boilers in the U.S. The major objectives of this work were to examine the impacts of the 1979 New Source Performance Standards on FGD system design and operation and to identify recent improvements in the technology. In the past four years, since the promulgation of the NSPS, the wet limestone process has been selected by utilities for 75 percent of the new plant capacity. In this time period, 77 plants representing over 37,000 MW of capacity have selected FGD systems. Several major trends in the design of limestone systems have become fairly standardized. Nearly all new systems are being built with spare absorber modules in order to qualify for the NSPS emergency by-pass provisions. The predominant absorber design is the open spray tower, due to minimal maintenance requirements. Forced oxidation to produce gypsum solids which can then be landfilled are two features being incorporated in many new units. Selection of the spray drying process for 15 percent of the new sites has also occurred in the last four years. The remaining throwaway and regenerable systems have not experienced any significant increases in applications. Finally, organic acid addition has been successfully demonstrated on limestone systems to improve the $SO_2$ removal and the system reliability. Organic acid addition is currently being used at two sites to upgrade the performance of older limestone systems.

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## CONVERTING UNITS OF MEASURE

Environmental Protection Agency policy is to express all measurements in Agency documents in metric units. In this report, however, to avoid undue cost or lack of clarity, some English units are used. Conversion factors from English to metric units are given below:

| To Convert From | To | Multiply By |
|---|---|---|
| Btu | J | 1055 |
| scfm (60°F) | $Nm^3$/hr (0°C) | 1.61 |
| cfm | $m^3$/hr | 1.70 |
| °F | °C | (°F-32)/1.8 |
| ft | m | 0.305 |
| ft/hr | m/hr | 0.305 |
| ft/sec | m/sec | 0.305 |
| $ft^2$ | $m^2$ | 0.0929 |
| gal/kacf | $l/m^3$ | 0.134 |
| gpm | l/min | 3.79 |
| in. | cm | 2.54 |
| in. $H_2O$ | mm Hg | 1.87 |
| lb. | g | 454 |
| lb-moles | g-moles | 454 |
| psi | kPa | 6.895 |
| ton | metric ton | 0.907 |

# SECTION 1
# SUMMARY

This report examines the current status of flue gas desulfurization (FGD) systems applied to coal-fired utility boilers for control of sulfur dioxide ($SO_2$) emissions. This work was performed for the Industrial Environmental Research Laboratory in Research Triangle Park, North Carolina (IERL-RTP) of the Environmental Protection Agency (EPA), to document changes in the design of FGD systems which have occurred over the years, specifically in response to the requirements of the 1979 New Source Performance Standards (NSPS) for utility boilers. This information is useful to EPA Regional Offices in evaluating $SO_2$ control technology for new utility applications, to the Office of Air Quality Planning and Standards (OAQPS) in their review of the 1979 NSPS which will be performed in the near future, and to the Office of Research and Development in establishing research initiatives. This information is also useful to A&E's, vendors, utility companies, and others interested in FGD technology.

In large part the objectives of this study were to examine differences in the designs and applications of FGD systems applied to the plants subject to the 1971 and 1979 NSPS. The basic difference between the two standards is the inclusion of a minimum percent reduction requirement of $SO_2$ emissions in the new standard. The 1971 standard limited $SO_2$ emissions to 1.2 pounds of $SO_2$ per million Btu. The 1979 NSPS requires a minimum of 70 percent removal for all controlled emissions less than 0.6 pounds of $SO_2$ per million Btu. A sliding percentage removal scale is used as sulfur levels increase. The maximum emission limit under the new standard is also 1.2 pounds of $SO_2$ per million Btu. Figure 1 shows the removal requirement of the NSPS for different sulfur levels. The major effect of the new standard has been to

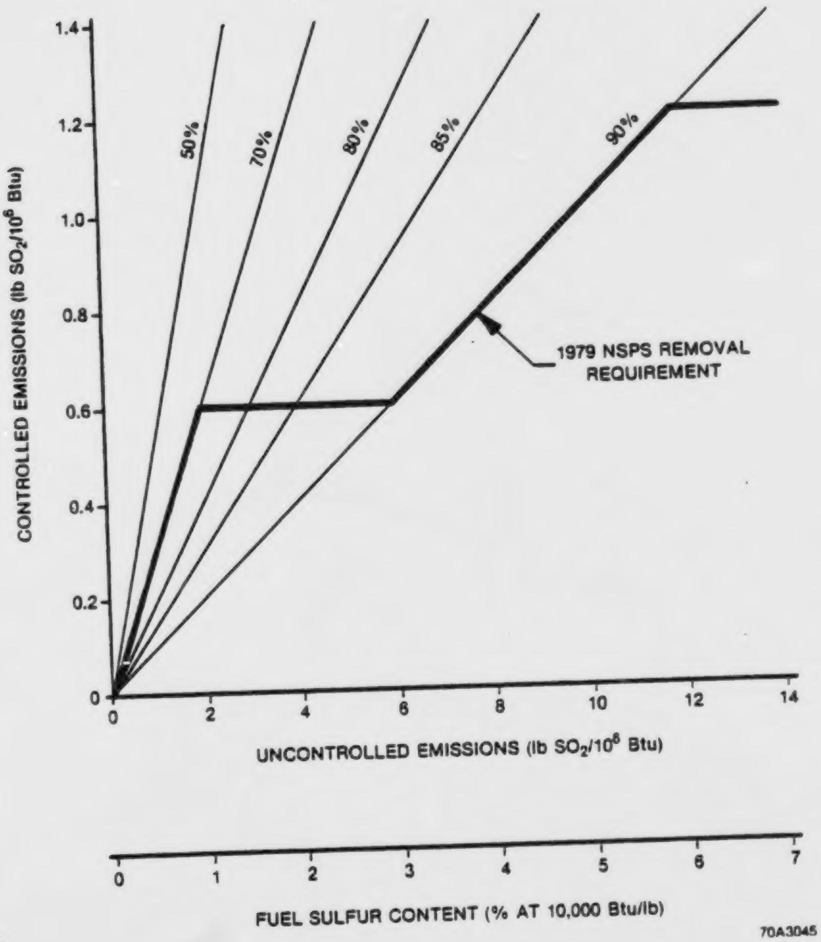

Figure 1. 1979 NSPS Removal Requirements

require $SO_2$ emission control on low sulfur coals which, under the 1971 standard, would not have required any $SO_2$ control. The rationale behind the decision to require $SO_2$ emission controls is discussed in detail in the Federal Register (5). Two major reasons cited were to promote the use of the best available FGD technology according to provisions of the Clean Air Act, and to maintain a competitive economic balance between high and low-sulfur coal producing regions in the U.S.

## FGD TECHNOLOGY STATUS

The most dominant trend in FGD technology is the overwhelming selection of the limestone FGD process for new units. Table 1 shows the status of the various FGD technologies as of September 1978, prior to promulgation of the 1979 NSPS, and as of January 1983. The equivalent scrubbed gas capacity (in megawatts) and number of units are presented for each process. The percentage distribution (by megawatts) of the various processes is also shown. The totals presented are for units operating, in construction, contracted, and planned which have selected an FGD technology. The EPA Utility FGD Survey, which is continually updated by PEDCo Environmental Inc. was used as a primary source of information for this table and others throughout this report.

The processes in Table 1 are listed in order of application over the last three years. As can be seen, limestone systems have been specified for seventy-five percent of the MW capacity of these units. Spray drying systems account for the majority of the remaining units. The other throwaway and regenerable technologies show only minor increases in application. The two processes showing decreases represent the shutdown of a magnesium oxide test unit and a reduction in size of several units which had planned (in 1978) to use alkaline ash scrubbing. Of particular note is the reduced use of the lime FGD process. In September 1978, 25 percent of the FGD capacity used the wet lime process. During the subsequent three year period, lime processes were only specified at 5 percent of the new units. Sixteen times as many limestone systems were selected as lime units in this interval. In 1978, the ratio was two limestone systems to each lime unit.

TABLE 1. COMPARISON OF FGD SYSTEM STATUS IN SEPTEMBER 1978 AND JANUARY 1983

| Process | September 1978[a] | | | January 1983[b] | | | Increase from 1978 to 1982 | | |
|---|---|---|---|---|---|---|---|---|---|
| | MW | Number | Percent (MW) | MW | Number | Percent (MW) | MW | Number | Percent (MW) |
| Limestone | 23,849 | 54 | 49.2 | 51,662 | 103 | 60.3 | 27,813 | 49 | 74.8 |
| Spray Drying | 400 | 1 | 0.8 | 6,353 | 18 | 7.4 | 5,953 | 17 | 16.0 |
| Lime | 12,065 | 29 | 24.9 | 13,736 | 32 | 16.0 | 1,671 | 3 | 4.5 |
| Sodium Carbonate | 884 | 4 | 1.8 | 3,155 | 9 | 3.7 | 2,271 | 5 | 6.1 |
| Dual Alkali | 1,102 | 3 | 2.3 | 2,288 | 6 | 2.7 | 1,186 | 3 | 3.2 |
| Wellman-Lord | 1,855 | 6 | 3.8 | 2,074 | 8 | 2.4 | 219 | 2 | 0.6 |
| Aqueous Carbonate | 100 | 1 | 0.2 | 100 | 1 | 0.1 | --- | -- | --- |
| Citrate | 60 | 1 | 0.1 | 60 | 1 | 0.1 | --- | -- | --- |
| Magnesium Oxide | 846 | 4 | 1.8 | 724 | 3 | 0.9 | <122> | <1> | <-0.3> |
| Alkaline Ash-Lime/Limestone | 7,316 | 14 | 15.1 | 5,493 | 13 | 6.4 | <1,823> | <1> | <-4.9> |
| Total | 48,477 | 117 | 100.0 | 85,645 | 194 | 100.0 | 37,168 | 77 | 100.0 |

[a] Reference 1

[b] Reference 2

The decline in the use of lime systems is due to three factors:

- The higher reagent cost of lime compared to limestone,
- The improvement in limestone system performance, and
- The development of spray drying technology for low sulfur applications.

The early limestone systems tended to have greater operating problems than lime systems. Additionally, the more reactive lime units achieved higher $SO_2$ removal efficiencies on the higher sulfur coals. Continued research by public and private interests has made the limestone system capable of meeting the current emission regulations for a wide variety of fuels. Consequently, the higher reagent cost of the lime process becomes a detrimental factor in process selection. Most economic studies show the lime process to be more economical than limestone systems only for small size units or low sulfur fuels. In both of these applications, the cost of lime is a fairly small portion of the total revenue requirement. However, the spray drying process has recently attained prominence in low sulfur applications, due to anticipated favorable economics and performance characteristics. Therefore, the relative ranking of the wet lime process will probably continue to decline in future applications. Lime is the predominant reagent used in the spray drying systems, which may result in an increase in the industry's demand for this reagent.

The regenerable systems (Wellman-Lord, magnesium oxide, Citrate, and Aqueous Carbonate) have not shown any substantial increase in applications. Part of the reason includes the developmental nature of two of these processes, as well as the higher costs associated with regenerable systems in general.

Table 2 shows the distribution of processes according to the emission requirements of each unit. Again, the totals are for systems operating, in construction, contracted, and planned. There are five emission regulation

### TABLE 2. FGD SYSTEMS AND REGULATORY CLASSIFICATIONS - JANUARY 1983

| Process | Applicable Emission Standard, MW (Number of Units) | | | | |
|---|---|---|---|---|---|
| | 1979 NSPS | 1971 NSPS | More Stringent Than 1979 NSPS | More Stringent Than 1971 NSPS, But Less Stringent Than 1979 NSPS | Less Stringent Than 1971 NSPS |
| Limestone | 27,461 (45) | 9,309 (25) | 7,239 (14) | 3,151 (10) | 4,502 (9) |
| Lime | 715 (2) | 4,551 (10) | 2,140 (4) | 5,486 (12) | 844 (4) |
| Spray Drying | 2,077 (8) | 1,180 (3) | 2,099 (4) | 887 (2) | 110 (1) |
| Alkaline Ash-Lime/Limestone | 475 (1) | 1,374 (4) | 1,979 (5) | 1,665 (3) | --- |
| Sodium Carbonate | 1,900 (4) | 375 (3) | 330 (1) | 550 (1) | --- |
| Wellman-Lord | --- | --- | 1,779 (4) | 295 (4) | --- |
| Dual Alkali | 1,107 (3) | 881 (2) | --- | 300 (1) | --- |
| Magnesium Oxide | --- | --- | --- | 724 (3) | --- |
| Aqueous Carbonate | --- | --- | --- | --- | --- |
| Citrate | --- | 60 (1) | --- | --- | 100 (1) |
| TOTAL | 33,735 (63) | 17,730 (48) | 15,566 (32) | 13,058 (36) | 5,556 (15) |

Source: Reference 2.

categories used in the FGD survey which have been defined by PEDCo. These include the 1971 and 1979 NSPS standards and state requirements which are more or less stringent than the federal standards.

Each of the processes in Table 2 except magnesium oxide, Aqueous Carbonate and Citrate, will or have been used in commerical applications to meet the 1979 NSPS or a more stringent state standard. However, the selection of a process technology is very site specific; not all of the processes are applicable to every situation. The following section discusses briefly the range of applications of each process and presents some general comments on performance, energy requirements, environmental effects, and reliability.

## FGD APPLICATIONS AND CHARACTERISTICS

Although most of the commercial FGD processes could be used at many types of sites, generally the plants using the same process have some similar characteristics which favors the use of the selected process. This section discusses these trends and also presents general comments on energy requirements, environmental considerations, and reliability. Table 3 summarizes these comments.

### Applications and Recent Process Improvements

The limestone process is the predominant FGD system in use today because of several factors. First, limestone deposits are found in many parts of the country which leads to low transportation and reagent costs. Secondly, limestone systems have been in use longer than other types of processes. Consequently, vendors and utilities are familar with the technology. Perhaps most importantly, until the advent of spray drying, limestone units have typically been the lowest cost process available. Commerical experience has also shown that limestone systems can be used on a wide variety of fuels successfully. Several innovations in process design have become commonplace in limestone process designs. The absorber vessel is almost always an open spray tower, which has lower maintenance requirements than packed towers. Forced oxidation of the sludge is being performed to produce better

TABLE 3. QUALITATIVE EVALUATION OF COMMERCIAL FGD PROCESSES

| Process | Performance | Energy Requirements | Environmental Considerations | Reliability |
|---|---|---|---|---|
| Limestone | New Systems are being designed to meet the 1979 NSPS with high and low sulfur coals. Additives can be used to enhance $SO_2$ removal. | Moderate to high. Depends on absorber pressure drop, amount of grinding required to prepare limestone. | Produces large volumes of sludge. | Historically the lowest of commercial processes, additives may improve performance. |
| Lime | Traditionally lime systems are somewhat more effective than limestone units at equivalent operating parameters. | Moderate, since reagent preparation does not require grinding. | Produces large volumes of sludge. | Better than limestone, however, still can experience scaling and plugging. |
| Spray Drying | Currently only able to meet NSPS for lower sulfur fuels. | Very low because of small pumping requirements and no reheat requirements. | Dry waste product often requires wetting for disposal. | Insufficient data base, however, process simplicity is an advantage. |
| Alkaline Ash | Limited to certain fuel types, usually only low sulfur coals. | Comparable to lime systems. | Produces large volume of flyash-FGD sludge. | Comparable to lime systems. |
| Sodium Carbonate | High $SO_2$ removal achievable. | Low, due to high alkalinity of reagent. | Large volume of soluble waste produced. Requires special disposal practices. | Very reliable due to simplicity. |
| Dual Alkali | High $SO_2$ removal achievable on high sulfur fuels. | Low due to high alkalinity of reagent. | Large volume of sludge produced. | Based on operating units, most reliable commercial system. |
| Wellman-Lord | High $SO_2$ removal achievable on high sulfur coal. | Fairly high because of steam requirements and gas for regeneration. | Produces sulfur, sulfuric acid, and sodium sulfate saleable by-products. | Fairly good, but may suffer somewhat due to process complexity. |
| Magnesium Oxide | High $SO_2$ removal achievable. | Moderate, coal can be used in drying and regeneration. | Produces sulfuric acid. | No commerical systems. Prototype plants had numerous mechanical-type problems. |

handling solids and reduce the scaling potential of the slurry. Additives such as magnesium and organic acids are being used commercially to raise the $SO_2$ removal closer to levels typically reached by more reactive sodium and lime systems. This flexibility has resulted in the large number of limestone systems in use. However, because of site specific factors and utility preferences, other processes have often been selected.

The most significant determinant of process selection is probably the fuel composition, specifically sulfur level, although the chloride and ash levels are also important. The sodium based processes (dual alkali, Wellman-Lord) are good candidates for high sulfur fuels because of the high sorption capability of the scrubbing liquor. One of the dual alkali system vendors is now marketing a process using limestone rather than the more expensive lime as the regenerant to improve process economics.

For low sulfur applications, spray drying is currently the most popular process, due to the reduced reagent volumes needed and reduced capital costs. Depending on the ash composition, the alkaline ash process can also be used in low sulfur applications, although no new units are being planned for this process. High chloride coals can cause both process chemistry problems and corrosion. A prescrubber before the $SO_2$ absorber is often used to remove the chloride, although prescrubber corrosion problems can result from the low pH liquors that are produced.

Land availability favors the regenerable over the throwaway processes such as the Wellman-Lord and magnesium oxide systems. These processes produce saleable by-products rather than throwaway sludges. The local availability of a reagent also affects the process selection. All sodium carbonate systems are near natural deposits of the material. If transportation costs are a major portion of the delivered reagent price, lime can have an advantage over limestone because of its greater reactivity per pound.

Water availability is often cited as a factor in FGD process selection, however, all of the processes use nearly the same amount of water. Spray drying may use slightly less water because the flue gas is not totally saturated; however, the dry waste product is moistened before disposal. The water requirements of the FGD system are typically less than 10 percent of the plant's total requirement. A more important consideration is the type of water used in the FGD system (3). The throwaway systems can generally tolerate poorer quality water than the regenerable systems.

## Energy Requirements

Energy requirements for FGD systems can vary significantly, depending upon the bases chosen for analysis. For example, the electrical and steam requirements for the commercial processes have been generally shown in previous reports to increase as follows, (lowest to highest); spray drying, sodium carbonate, dual alkali, alkaline ash, lime, limestone, magnesium oxide, and Wellman-Lord. The energy requirement as a percentage of new power plant output is often reported to be less than one percent for the sodium systems, normally about 3 percent for limestone systems (excluding reheat) and even 10 percent for some regenerable units. The problem with comparing the energy consumption required by the FGD system alone is that it does not reflect the energy required to process the reagents or the sludge, nor does it include any by-product energy credit. These energy requirements are reflected in the reagent cost and by-product credits.

TVA performed a study which examined the ground-to-ground energy requirements of three FGD systems (4). The design basis was a 500 MW unit burning 3.5 percent sulfur coal meeting the 1971 NSPS. The results for three FGD processes are shown in Table 4. The lime system had the lowest energy requirement if only the FGD system is considered. However, the energy required to calcine limestone into lime (which is reflected in the price of lime) is significant. When all of these energy requirements are considered, the limestone process has the lowest energy requirement. The credit for sulfuric acid in the magnesium oxide process reflects the energy required to mine and convert an equivalent amount of elemental sulfur. This credit makes the regenerable process almost as efficient as the lime system. Since spray drying and dual alkali systems use lime, their overall ground-to-ground energy requirements are much closer to the other systems than would be indicated by just comparing the FGD system energy requirement.

Table 4. FGD GROUND-TO-GROUND ENERGY REQUIREMENTS

| Function | Btu/lb Sulfur Removed | | |
|---|---|---|---|
| | Limestone | Lime | Magnesium Oxide |
| Mining | 438 | 356 | 25 |
| Absorbent Processing | - | 6,198 | 161 |
| Transportation | 176 | 143 | 33 |
| FGD | 14,042 | 13,165 | 26,387 |
| Waste Disposal | 22 | 15 | - |
| By-product Credit | - | - | (5,491) |
| Total | 14,678 | 19,877 | 21,115 |
| Btu/KWh | 291 | 395 | 420 |
| % Gross Power Unit Output | 3.2 | 4.4 | 4.7 |

Source: Reference 4

## Environmental Considerations

All six of the commercial throwaway FGD systems produce large volumes of waste material requiring disposal. The calcium based sludge from the limestone, lime, alkaline ash, and dual alkali processes are very similar in composition. Typically, this sludge is disposed of in a clay lined pond. A thickened slurry is pumped to the pond where the solids settle out. Supernatant water is usually returned to the FGD system. Many of the newer plants are using landfilling for disposal. In this practice, the sludge is vacuum filtered or centrifuged to a higher solids content, and mixed with lime and/or fly ash to form a stable mass. In either installation, some type of liner material is used to prevent ground water contamination due to leachate from the sludge.

Solid waste from spray drying, if lime is used as the reagent, is also similar to the material described above, except that it is a dry powder. However, in order to dispose of this waste, it is often wetted to increase its compressive strength and to reduce dusting problems. Several research programs are being sponsored by the EPA and EPRI to determine the best methods of disposing of spray drying wastes.

The sodium carbonate process produces a liquid purge stream high in sodium sulfites and sulfates. The plants which use the process, use evaporation ponds to crystallize the salts. This technique is only applicable in arid regions. Pond liners may also be necessary to prevent ground water contamination.

The regenerable processes have a much smaller secondary environmental impact since they produce saleable by-products which are sold to user industries. However, lime and limestone processes may be designed to produce gypsum wallboard or other by-products if desired.

## Reliability

The reliability of an FGD system is a major concern to utilities operating plants subject to the 1979 NSPS. Older plants and regulations often allowed bypassing flue gas around malfunctioning absorber modules or the entire system. However, the 1979 NSPS has significantly reduced the possibility of by passing the FGD system except in emergency situations. "The final standards allow an owner or operator to bypass uncontrolled flue gas around a malfunctioning FGD system provided, (1) the FGD system has been constructed with a spare FGD module, (2) FGD modules are not available in sufficient numbers to treat the entire quantity of flue gas generated, and (3) all available electric generating capacity is being utilized in a power pool or network consisting of the generating capacity of the affected utility company (except for the capacity of the largest single generating unit in the company), and the amount of power that could be purchased from neighboring interconnected utility companies(5)." From the constraints expressed in the third requirement it is evident that utility companies are expected to install reliable FGD equipment. This requirement also places an economic incentive on the utilities to obtain reliable operation from their FGD systems, since the available electric generating capacity will often include uneconomic gas-fired turbines and oil or gas-fired boilers.

The few units operating regulated by the 1979 NSPS have not experienced sufficient operating time to determine if they are more reliable than older units. Several studies have addressed the reliability of some of the older units. In a study by the National Electric Reliability Council (NERC), the causes of outages for FGD systems of 400 MW and larger have been categorized (6). The results of this study are presented in Table 5. As can be seen, two-thirds of the outage time was caused by non-specific FGD technology problems. This would seem to indicate that FGD processes often suffer a poor reliability image because of peripheral equipment failures rather than process related problems.

Table 5. PERCENT OF PLANT OUTAGE TIME CAUSED BY SPECIFIED FGD COMPONENT FAILURE (NERC DATA)

| Component | Percent of Outage Time |
|---|---|
| Dampers | 28 |
| Duct system, baffles | 19 |
| Absorber tower | 16 |
| Fans, ID, FD, blades, vibration | 17 |
| Mist eliminator | 9 |
| Reaction tank | 4 |
| Spray pumps | 3 |
| External causes | 2 |
| Thickener | 1 |
| Test programs | 1 |

The FGD data base maintained by PEDCo contains information on availability and operability of the individual plants. The average availability and operability of the commercial FGD processes are presented in Table 6, along with their definition of these two terms. The availability index can be artifically high if the boiler is not in operation although the FGD system could operate. The operability index is a more accurate estimate of the reliability of the various processes since it involves the actual time that the boiler is operated. It can be lowered, however, if a decision is made not to operate the FGD system.

This table indicates that limestone and lime systems are generally the most unreliable processes. However, it is difficult to compare reliability data of various processes because of differences in plant design and operating requirements. There is no data available from plants regulated by the 1979 NSPS. Very few of the operating systems in the PEDCo data base have spare modules which increase reliability. Additionally, the design differences make it difficult to compare reliability figures. New units are incorporating design improvements obtained from experiences with existing

Table 6. DEPENDABILITIES OF FGD SYSTEMS (MARCH 1981-MARCH 1982)

| Process | Number of Operating Systems | Systems Included In Averages | Availability[a] | Operability[b] |
|---|---|---|---|---|
| Limestone | 39 | 24 | 73.5 | 73.8 |
| Lime | 24 | 23 | 84.1 | 75.4 |
| Wellman-Lord | 7 | 7 | 88.5 | 70.0 |
| Alkaline Ash | 9 | 6 | 89.3 | 78.0 |
| Sodium Carbonate | 5 | 4 | 89.9 | 87.4 |
| Dual Alkali | 3 | 3 | 96.2 | 79.7 |

[a] Percentage of hours the FGD system is available (whether used or not) for operation divided by the hours in the period.

[b] Percentage of hours the FGD system was operated divided by the boiler operating hours during the period.

Source: Reference 7

systems. Many of the older limestone and lime systems with poor designs are included in the averages. Their recurring problems reduce the reliability figures for the process. Consequently, with the construction of new units there should be a gradual increase in the overall process reliability. The effects of a new technology, such as the use of adipic acid with limestone systems, can also significantly improve the reliability of a system. However, it is difficult to quantify in advance the magnitude of reliability increases due to process improvements.

One recent study examined the effect of forced outages of scrubber modules on utility system production costs (8). Two different utility fuel-mix systems were analyzed. The results indicate that in many instances, the addition of a second spare module is economically justified. For the system based primarily on coal capacity (49 percent coal, 24 percent oil), module outage time greater than 14 percent favors using a second spare. In the oil based system (19 percent coal, 57 percent oil), two spare modules are justified above a 10 percent module forced outage rate. The assumptions used in this study are site-specific, however, they do provide an interesting perspective on the reliability issue. However, the most crucial piece of data, the actual module outage rate for a well designed system, is not given.

## ECONOMICS

The costs of FGD systems are very site specific. The coal composition, degree of control required, and economic bases all have a major influence on the reported or estimated costs. Consequently, the actual cost of a "standard" FGD system can vary considerably, which leads to some confusion as to what the costs actually are. Several recent reports have estimated the costs of various FGD processes for a standard size plant by examining the various design and economic bases. TVA has published the most recent economic study of limestone and lime systems which uses a mid-1982 time basis for capital costs. The standard plant is a 500 MW unit burning 3.5 percent sulfur bituminous coal meeting the 1979 NSPS. A bypass duct capable of handling 50 percent of the flue gas is also included. Although there are many site specific factors which affect the capital cost of an FGD system, four parameters account for the majority of the cost differences between estimates for a standard size plant and fuel. These parameters are the estimate date, use of a spare module, bypass ducts, and the $SO_2$ removal requirement.

By examining the various site specific cost factors used to estimate the costs of several FGD systems, and adjusting the factors to a common basis, a fairly consistent set of cost estimates can be obtained. The results of this exercise are presented in Table 7 and illustrated in Figures 2 and 3. As presented in Table 7 and Figure 2, the "raw" or unadjusted cost estimates reported for each of six commercial FGD processes vary considerably. However, when these cost estimates are adjusted to a common basis (Figure 3) the variance in the estimate significantly decreases.

The cost adjustments were made by using the Chemical Engineering plant cost index to adjust for inflation. A spare module and associated equipment for a 500 MW unit was assumed to increase capital costs 25 percent. Bypass ducts increased the costs five and ten percent respectively for 50 and 100 percent bypass. Increasing the $SO_2$ removal from the 1971 NSPS to the 1979 NSPS was estimated as a five percent increase.

As shown in Table 7, the 1982 cost of most of the FGD systems is about $200 per KW. The average costs show the same relative rankings as in individual reports which estimate several processes. Lime systems have the lowest capital cost, followed by limestone and dual alkali processes. The regenerable systems are 20-30 percent higher. The spray drying systems include integral particulate control, and have an average cost of $208 per KW. ESP's for the other processes will add about 20 percent to the total air pollution control equipment cost. It should be noted that the designs of the spray drying units have not been successfully demonstrated for a high-sulfur coal.

TABLE 7. ESTIMATES OF FGD SYSTEM COSTS

Limestone Systems

| Reference | 9 | 10 | 11 | 12 | 13 | Average |
|---|---|---|---|---|---|---|
| Cost ($/KW) | 206 | 157 | 165 | 186 | 98 | 162 |
| Year | 1982 | 1978 | 1979 | 1978 | 1979 | |
| Inflation Factor | 1.0 | 1.36 | 1.25 | 1.36 | 1.25 | |
| Spare Module Factor | 1.0 | 1.0 | 1.0 | 1.0 | 1.25 | |
| Bypass Duct Factor | 1.0 | 0.95 | 0.95 | 0.95 | 1.05 | |
| $SO_2$ Removal Factor | 1.0 | 1.0 | 1.0 | 1.0 | 1.05 | |
| 1982 Cost ($/KW) | 206 | 203 | 195 | 240 | 169 | 203 |

Lime Systems

| Reference | 9 | 10 | 4 | Average |
|---|---|---|---|---|
| Cost ($/KW) | 192 | 149 | 90 | 144 |
| Year | 1982 | 1978 | 1979 | |
| Inflation Factor | 1.0 | 1.36 | 1.25 | |
| Spare Module Factor | 1.0 | 1.0 | 1.25 | |
| Bypass Duct Factor | 1.0 | 0.95 | 1.05 | |
| $SO_2$ Removal Factor | 1.0 | 1.0 | 1.05 | |
| 1982 Cost ($/KW) | 192 | 193 | 155 | 180 |

Dual Alkali Systems

| Reference | 10 | 13 | Average |
|---|---|---|---|
| Cost ($/KW) | 166 | 101 | 134 |
| Year | 1978 | 1979 | |
| Inflation Factor | 1.36 | 1.25 | |
| Spare Module Factor | 1.0 | 1.25 | |
| Bypass Duct Factor | 0.95 | 1.05 | |
| $SO_2$ Removal Factor | 1.0 | 1.05 | |
| 1982 Cost ($/KW) | 216 | 174 | 195 |

(Continued)

TABLE 7. ESTIMATES OF FGD SYSTEM COSTS (Continued)

Spray Drying Systems

| Reference | 14 | 15 | 10 | Average |
|---|---|---|---|---|
| Cost ($/KW) | 200 | 102 | 192 | 165 |
| Year | 1982 | 1979 | 1978 | |
| Inflation Factor | 1.0 | 1.25 | 1.36 | |
| Spare Module Factor | 1.0 | 1.25 | 1.0 | |
| Bypass Duct Factor | 1.0 | 1.05 | 0.95 | |
| $SO_2$ Removal Factor | 1.0 | 1.05 | 1.0 | |
| 1982 Cost ($/KW) | 200 | 176 | 248 | 208 |

Wellman-Lord Systems

| Reference | 10 | 16 | 17 | 12 | Average |
|---|---|---|---|---|---|
| Cost ($/KW) | 180 | 131 | 126 | 267 | 176 |
| Year | 1978 | 1979 | 1977 | 1978 | |
| Inflation Factor | 1.36 | 1.25 | 1.51 | 1.36 | |
| Spare Module Factor | 1.0 | 1.25 | 1.0 | 1.0 | |
| Bypass Duct Factor | 0.95 | 1.05 | 0.95 | 0.95 | |
| $SO_2$ Removal Factor | 1.0 | 1.05 | 1.0 | 1.0 | |
| 1982 Cost ($/KW) | 232 | 225 | 180 | 345 | 246 |

Magnesium Oxide Systems

| Reference | 10 | 4 | 17 | 11 | Average |
|---|---|---|---|---|---|
| Cost ($/KW) | 176 | 132 | 136 | 193 | 159 |
| Year | 1978 | 1979 | 1977 | 1979 | |
| Inflation Factor | 1.36 | 1.25 | 1.51 | 1.25 | |
| Spare Module Factor | 1.0 | 1.25 | 1.0 | 1.0 | |
| Bypass Duct Factor | 0.95 | 1.05 | 0.95 | 0.95 | |
| $SO_2$ Removal Factor | 1.0 | 1.05 | 1.0 | 1.0 | |
| 1982 Cost ($/KW) | 228 | 228 | 194 | 230 | 220 |

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[a] Includes baghouse for particulate control. ESP for other FGD systems estimated at $38/KW in 1982 dollars which is not included in the process cost.

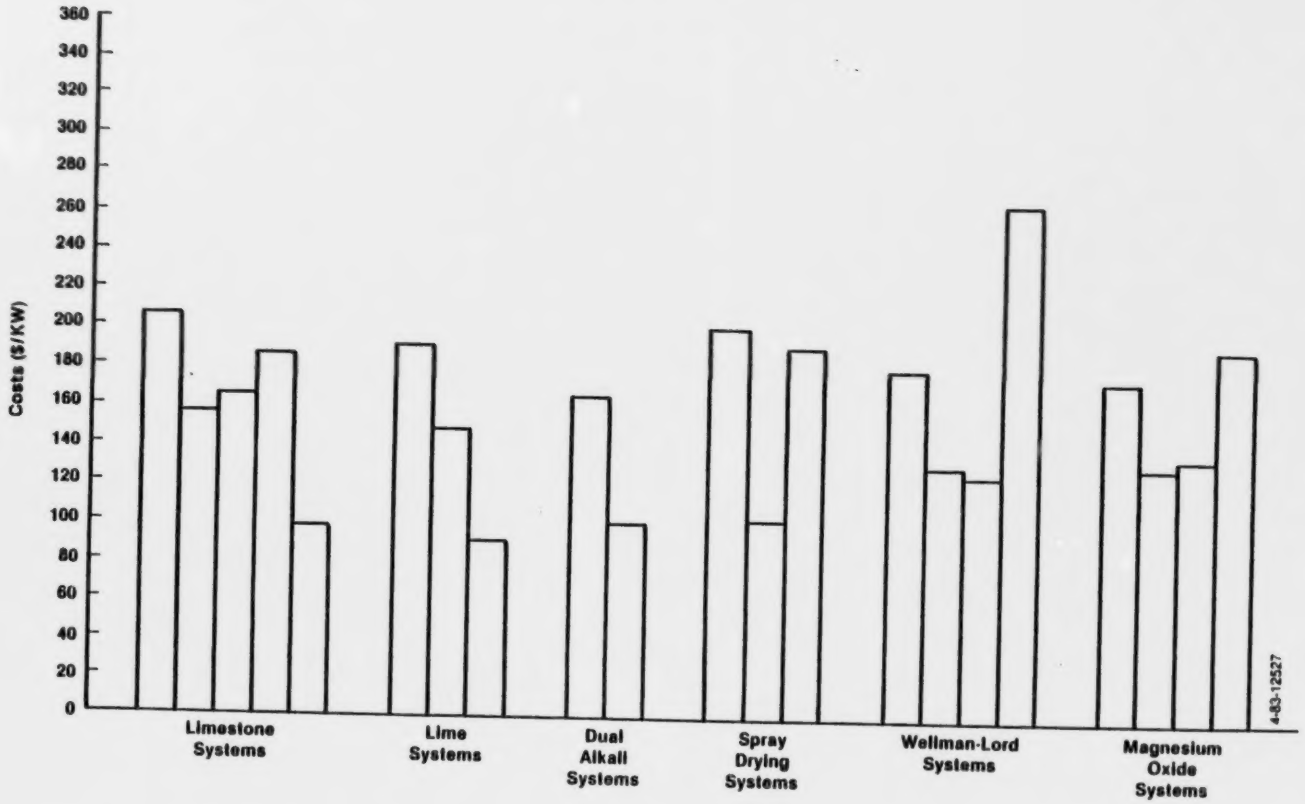

Figure 2. Raw Estimates of FGD System Costs

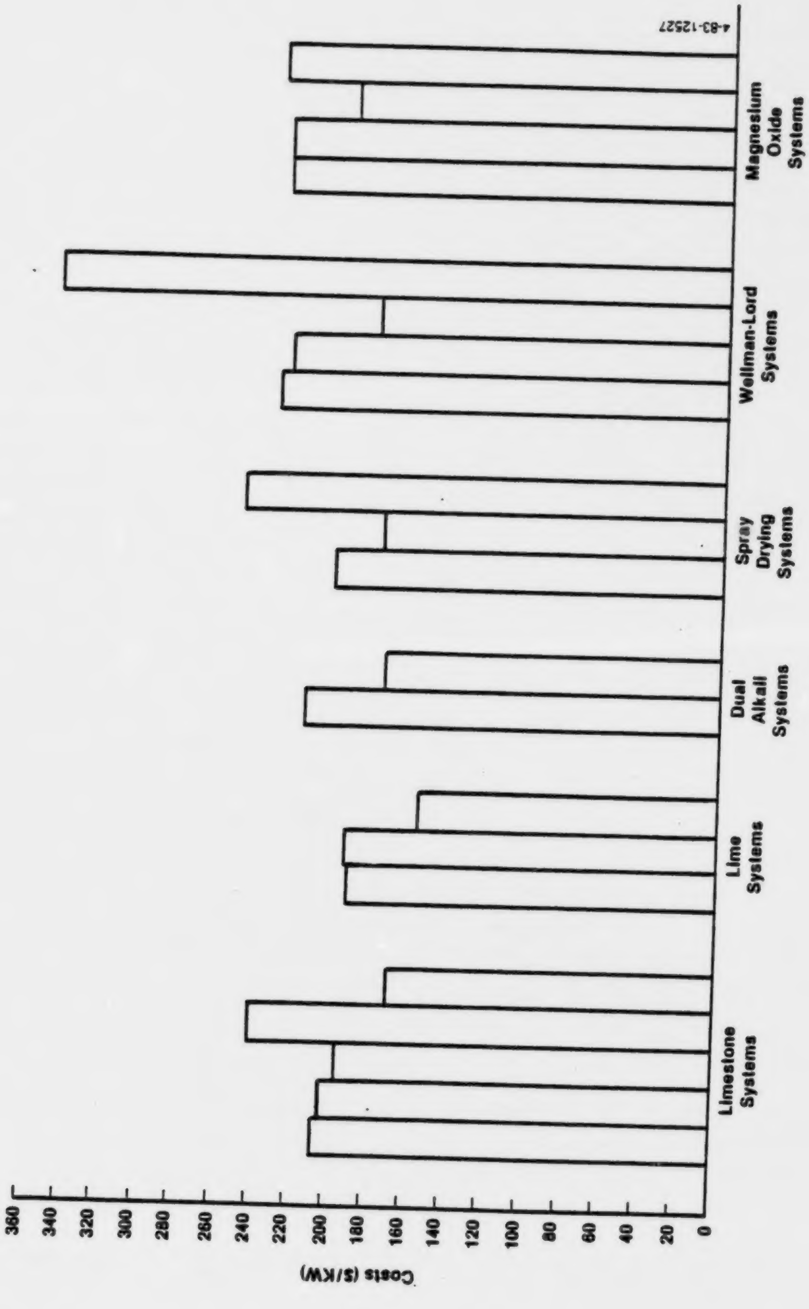

Figure 3. FGD System Cost Estimates Adjusted to Consistent Basis

The annual costs of the different FGD processes depend as much on capital charges as fixed and variable operating costs. There are too many variables (reagent costs, interest rates, labor rates, operating profile return on investment (ROI), etc.) to use a simplistic approach to present standarized annual costs as was done for the capital costs. In individual reports using the same bases, limestone and spray drying are usually the least expensive followed by lime and dual alkali, and then, the regener-able systems.

The utility FGD survey performed by PEDCo compiles reported cost data on operating systems and adjusts the costs to a common basis. Averages of the different processes are shown in Table 8. The low sulfur systems have both the lowest capital and operating costs. The high sulfur limestone, lime, and dual alkali units all cost between $140-150 per KW. Since these are operating systems, they are predominantly units designed for the 1971

Table 8. AVERAGE COSTS OF OPERATING FGD SYSTEMS (1981 DOLLARS)

| Process | Number of Units | Capital Cost ($/KW) | Annual Cost (mills/KWh) |
|---|---|---|---|
| Limestone-Low Sulfur[a] | 16 | 79 | 4.8 |
| Lime-Low Sulfur | 5 | 72 | 5.1 |
| Sodium Carbonate | 4 | 111 | 6.4 |
| Alkaline Ash | 7 | 118 | 6.7 |
| Dual Alkali | 3 | 147 | 8.7 |
| Limestone-High Sulfur[b] | 7 | 149 | 9.2 |
| Lime-High Sulfur | 15 | 141 | 10.5 |
| Wellman-Lord | 3 | 272 | 18.1 |

[a] Low sulfur - <1.5 percent

[b] High sulfur - >1.5 percent

Source: Reference 2

NSPS, with no spare modules or bypass ducts and a lower $SO_2$ removal requirement. Using the previous cost multipliers results in a 1982 cost of $200-220 per KW, which is very close to the estimated costs previously presented. As can be seen, when the costs are placed on a common basis, they agree very well.

## $SO_2$ REMOVAL TEST RESULTS

Prior to promulgation of the 1979 NSPS, the EPA performed emission tests on various types of FGD systems. These units were built to meet the 1971 NSPS or state standards, which were not as stringent as the current NSPS. However, many of the units tested meet the 1979 NSPS, or are very close to satisfying its requirements. Based on these test results, the EPA administrator set the new standard at a level attainable by currently available technology. The list of sites tested and the test results are presented in Table 9. In addition, a comment column is provided to indicate the $SO_2$ removal level that would be required for each of the sites to achieve compliance with the 1979 NSPS. As can be seen, seven of the thirteen units built under more lenient standards meet the current NSPS. Several of the other units were within a few percentage points of meeting the standard. Consequently, designing and operating an FGD system to meet the 1979 NSPS required only a moderate advance in the status of FGD technology.

Two deficiencies are readily apparent in the continuous emission monitoring results data base which need to be recognized. The limestone systems tested were not typical of the systems currently being designed and built which are predominantly using spray towers. Secondly, the growing acceptance of spray drying by the utility industry indicates that a number of these units should also be tested.

## CONCLUSIONS

The 1979 New Source Performance Standards for utility boilers requires flue gas desulfurization systems on all new coal-fired boilers. Since its initial promulgation on September 18, 1978, 73 FGD systems representing over

TABLE 9. FGD SYSTEM CONTINUOUS EMISSION MONITORING RESULTS

| Process | Unit | $SO_2$ Removal Efficiency (percent) | Emission Rate (lbs $SO_2/10^6$ Btu) | Comments |
|---|---|---|---|---|
| Limestone | Shawnee V/ST[a] | 96.1 | 0.22 | meets 1979 NSPS |
| | Lawrence 4 | 96.6 | 0.03 | meets 1979 NSPS |
| | Southwest 1 | >90 | <0.64 | meets 1979 NSPS |
| Lime | Cane Run 4 | 83.8 | 0.88 | 88.8% Removal needed for 1979 NSPS |
| | Conesville 5 | 89.2 | 0.80 | 90.0% Removal needed for 1979 NSPS |
| | Bruce Mansfield 1 | 85.3 | 0.78 | 88.5% Removal needed for 1979 NSPS |
| | Shawnee TCA | 88.6 | 0.64 | 89.3% Removal needed for 1979 NSPS |
| | Pleasants 1 | >90 | <0.50 | meets 1979 NSPS |
| Dual Alkali | A.B. Brown 1 | 87.7 | 0.85 | 90.0% Removal needed for 1979 NSPS |
| | Cane Run 6 | 91.6 | 0.70 | meets 1979 NSPS |
| Wellman-Lord | D.H. Mitchell 11 | 89.6 | 0.60 | meets 1979 NSPS |
| Magnesium Oxide | Eddystone 1A | 96.8 | 0.16 | meets 1979 NSPS |
| Spray Drying | Celanese Amcelle[b] | 70.0 | 0.92 | 80.4% Removal needed for 1979 NSPS |

[a] Units tested with adipic acid additdive

[b] Industrial boiler, utility $SO_2$ NSPS not directly comparable

35,000 MW of capacity have been constructed, contracted, or planned by the utility industry. In the previous seven years, 48,000 MW (117 units) had been built with FGD systems. Since the promulgation of the 1979 NSPS, the wet limestone FGD process has been selected for nearly two-thirds of the future plants. The predominance of this process is due to its technical flexibility and favorable economics. Improvements in reliability and efficiency due to the use of spray towers, forced oxidation, and $SO_2$ removal enhancement additives enable the limestone process to be used in many applications.

The use of spray drying to control $SO_2$ and particulates from lower sulfur coal is the only other technology to be widely favored in the industry. Many commerical spray drying systems will soon be on-line demonstrating this technology. The other throwaway systems (lime, dual alkali, alkaline ash, sodium carbonate) will be used infrequently at new plants. Higher reagent costs are the primary reason for the lack of new orders for these systems. The regenerable systems, due to their higher costs, will only be used in specialized applications, notably land restricted sites, and those with favorable $SO_2$ regeneration opportunities.

Results of this study indicate that the most predominant FGD processes in recent use by the utility industry for $SO_2$ control are the limestone and spray drying processes. Therefore, future research efforts to improve the operations of limestone systems and to develop a better understanding of the spray drying process will produce results that are of most benefit to the utility industry. With regard to the limestone process, additional research efforts to focus on the following areas is needed in order to solve existing problems, increase on-stream operating time, and decrease process operating costs:

1) Improvements in $SO_2$ monitors and pH controllers.

2) Investigation of new materials to reduce corrosion.

3) Optimization of the limestone preparation area of the process.

4) Additional improvements in the area of sludge disposal particularly in the areas of leaching control, improved oxidation techniques, and alternate end uses.

The major research need for the spray drying process is to develop a fundamental understanding of its reaction mechanisms. A more fundamental understanding of this process could perhaps extend its use to high sulfur coals and result in process improvements which could lead to lower process costs.

## SECTION 2

## CURRENT STATUS

This section discusses the current status of commercial utility flue gas desulfurization systems in operation and under construction. First, a comparison of the number and types of systems as of September 1978 (prior to proposal of the revised NSPS) and of January 1983 is given. A brief discussion of the general effects of the revised NSPS is presented, followed by a discussion of the individual FGD technologies with emphasis on design changes and recent process improvements. Finally, emerging FGD technologies which have been tested or are undergoing further development are briefly discussed.

### OVERALL STATUS

The EPA has sponsored a survey of utility FGD systems for over eight years. This project, which is conducted by PEDCo Environmental Incorporated, monitors the status of commercial-scale FGD systems applied to coal-fired utility boilers. Information voluntarily supplied by utilities, vendors, researchers, and regulatory agencies is stored in a computerized data base. In an annual report, with quarterly updates, both design and performance data are presented for operating FGD systems. In addition, the FGD systems are classified by their regulatory status and comments are presented on their overall operation.

Table 10 presents the operational status and types of FGD systems as of September 1978, which was just prior to the issuance of the revised NSPS. The processes are listed in order of total megawatts of flue gas

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## TABLE 10. STATUS OF FGD SYSTEMS - SEPTEMBER 1978

| Process | MW Capacity (number of units) | | | | |
|---|---|---|---|---|---|
| | Operating | In Construction | Contracted | Planned | Total |
| Limestone | 7,334 (20) | 8,497 (20) | 3,968 (8) | 4,050 (6) | 23,849 (54) |
| Lime | 4,900 (14) | 4,880 (10) | 2,285 (5) | - | 12,065 (29) |
| Alkaline Ash-Lime/Limestone | 2,590 (5) | 500 (1) | 3,647 (5) | 579 (3) | 7,316 (14) |
| Wellman-Lord | 735 (3) | 180 (1) | 940 (2) | - | 1,855 (6) |
| Dual Alkali | - | 1,102 (3) | - | - | 1,102 (3) |
| Sodium Carbonate | 375 (3) | 509 (1) | - | - | 884 (4) |
| Magnesium Oxide | 120 (1) | - | - | 726 (3) | 846 (4) |
| Spray Drying | - | 400 (1) | - | - | 400 (1) |
| Aqueous Carbonate | - | - | 100 (1) | - | 100 (1) |
| Citrate | - | 60 (1) | - | - | 60 (1) |
| TOTAL | 16,054 (46) | 16,128 (38) | 10,940 (21) | 5,355 (12) | 48,849 (117) |

Source: Reference 1.

scrubbed. As can be seen, limestone, lime, and alkaline ash systems using lime or limestone accounted for over 89 percent of the total FGD capacity in use or planned in 1978.

The current NSPS applies to any steam generating units for which construction was commenced after September 19, 1978. Of the 117 units listed in Table 10, five had not begun construction when the revised NSPS was proposed. These units were therefore subject to the revised standard.

Table 11 lists the status of FGD systems as of January 1983, again in decreasing order of megawatts scrubbed. An additional 77 units are identified in this table, the majority of which are subject to the 1979 NSPS or more stringent state standards. Some of the units are retrofits and are regulated by less stringent standards. Limestone, lime, and alkaline ash systems account for 80 percent of the megawatts scrubbed. The most striking difference observed in Table 11 is the number of spray drying systems. In Table 10, only one spray drying system was listed.

## GENERAL NSPS REQUIREMENTS

The initial NSPS promulgated on December 23, 1971, for utility boilers limited $SO_2$ emissions to 1.2 pounds of $SO_2$ per million Btu's of heat input (18). Most high Btu coals with less than one percent sulfur could be burned in compliance without using an FGD system. Slightly higher sulfur coals required partial scrubbing in order to meet the standard. Only coals with six percent sulfur or higher required 90 percent $SO_2$ control (for a heat content of 10,000 Btu per pound).

The current utility NSPS, revised in 1979, limits $SO_2$ emissions to a maximum of 1.2 pounds per million Btu heat input and a 90 percent reduction based on a 30-day rolling average (5). A minimum 70 percent reduction is required for emission rates of less than 0.6 pounds per million Btu. Intermediate or low sulfur coals are permitted a variable removal efficiency between 70 and 90 percent provided their emissions are less than 0.6 pounds per million Btu. Coals that have a sulfur content of less than about one

TABLE 11. STATUS OF FGD SYSTEMS - JANUARY 1983

| Process | MW Capacity (number of systems) | | | | |
|---|---|---|---|---|---|
| | Operating | In Construction | Contracted | Planned | Total |
| Limestone | 14,118 (41) | 8,179 (18) | 7,225 (11) | 22,140 (33) | 51,662 (103) |
| Lime | 9,886 (24) | 2,870 (6) | 650 (1) | 330 (1) | 13,736 (32) |
| Spray Drying | 550 | 2,438 (8) | 2,079 (5) | 1,836 (3) | 6,353 (18) |
| Alkaline Ash Lime/Limestone | 4,093 | 1,400 (2) | - | - | 5,493 (13) |
| Sodium Carbonate | 1,225 (5) | - | - | 1,900 (4) | 3,155 (9) |
| Dual Alkali | 1,181 (3) | 421 (1) | 686 (2) | - | 2,288 (6) |
| Wellman-Lord | 1,540 (7) | 534 (1) | - | - | 2,074 (8) |
| Magnesium Oxide | - | 724 (3) | - | - | 724 (3) |
| Aqueous Carbonate | - | 100 (1) | - | - | 100 (1) |
| Citrate | 60 (1) | - | - | - | 60 (1) |
| TOTAL | 32,683 (94) | 16,666 (4) | 10,640 (19) | 25,656 (41) | 85,645 (194) |

Source: Reference 2.

percent require only 70 percent removal, while coals with sulfur percentages of six percent or higher require a mean removal efficiency above 90 percent. A 90 percent, 30-day average corresponds to a 92 to 94 mean efficiency depending on the variability of the individual FGD system (21).

Table 12 illustrates the number of FGD systems affected by the 1971 and 1979 NSPS by showing the number and capacity of coal-fired generating units built since 1970. This table can be divided into four time periods which roughly correspond to the different $SO_2$ emission regulations. The NSPS, as its name implies, is generally only applicable to new sources of emissions. Both the 1971 and 1979 standards exempted units under construction from meeting the standard. As a result, until 1975, no units were completed which were regulated by the 1971 standard. It should be noted that the start-up date in Table 12 is for the boiler and generator. Many of the initial FGD systems were retrofit to older boilers to obtain operating experience and are not included in Table 12. Additionally, the ten FGD systems shown in the period from 1970 to 1975 did not necessarily commence operation when their respective units started.

From the period of 1976 to 1984, plants were and are being built in accordance with the 1971 standard which allows the use of low-sulfur coal to achieve compliance. As can be seen, slightly more plants were built without FGD systems than with (102 to 90) during this period. The 1979 NSPS affects plants being built in the 1985-1990 time frame and resulted in a significant shift to FGD systems for new capacity. Some of the units in this period are still regulated by the 1971 NSPS, however most of the units listed without FGD systems have not as of yet selected a technology. Some of these units may also have been cancelled or delayed due to a reduced growth in electricity requirements. The total number of units which will be built with or without FGD systems after 1990 is unknown due to uncertainties in load requirements, economic climate, and regulatory standards.

Another important aspect of the 1979 NSPS affecting the design of FGD systems involves the bypass provisions. The standard allows an operator to bypass uncontrolled flue gases around a malfunctioning FGD system provided:

TABLE 12. NEW COAL-FIRED UNITS WITH AND WITHOUT FGD SYSTEMS

| Unit Startup | Systems Without FGD | | Systems With FGD | |
|---|---|---|---|---|
| | MW | Number | MW | Number |
| 1970 | 11,599 | 27 | 115 | 1 |
| 1971 | 13,631 | 23 | 750 | 2 |
| 1972 | 11,756 | 22 | 356 | 1 |
| 1973 | 13,422 | 22 | 1,203 | 2 |
| 1974 | 9,711 | 16 | 864 | 2 |
| 1975 | 8,016 | 13 | 866 | 2 |
| 1976 | 5,265 | 11 | 3,953 | 9 |
| 1977 | 6,511 | 11 | 4,836 | 8 |
| 1978 | 7,397 | 15 | 4,795 | 12 |
| 1979 | 6,948 | 13 | 4,377 | 10 |
| 1980 | 10,390 | 18 | 5,421 | 10 |
| 1981 | 4,707 | 12 | 4,975 | 11 |
| 1982 | 5,138 | 9 | 5,109 | 11 |
| 1983 | 2,770 | 4 | 4,306 | 11 |
| 1984 | 5,910 | 9 | 3,950 | 8 |
| 1985 | 3,622 | 5 | 12,571 | 23 |
| 1986 | 3,417 | 6 | 5,081 | 8 |
| 1987 | 1,751 | 3 | 9,898 | 16 |
| 1988 | 3,240 | 6 | 8,091 | 11 |
| 1989 | 4,551 | 7 | 7,907 | 12 |
| 1990 | 3,615 | 7 | 3,493 | 5 |
| 1991 | 1,760 | 3 | 708 | 1 |
| 1992 | 1,245 | 2 | 3,185 | 5 |
| 1993 | 100 | 1 | 1,890 | 2 |
| 1994 | – | – | 1,300 | 2 |
| 1995 | – | – | – | – |
| 1996 | – | – | 1,000 | 2 |
| TOTAL | 146,472 | 265 | 101,000 | 188 |

Source: Reference 22.

(1) the FGD system has been constructed with a spare FGD module, (2) FGD modules are not available in sufficient numbers to treat the entire quantity of flue gas generated, and (3) all available electric generating capacity is being utilized in a power pool or network consisting of the generating capacity of the affected utility company (except for the capacity of the largest single generating unit in the company), and the amount of power that could be purchased from neighboring interconnected utility companies has been purchased (5).

This regulation stipulates that in order to bypass flue gas around a malfunctioning FGD system, at least one spare module must be installed, and that a reduction in the unit load would be deleterious to the reliability of electric service. Consequently, most new FGD systems are now being designed with a spare module. Units with a generating capacity of 125 MW or less are exempt from the spare module provision since they usually only require one module and would incur a disproportionate cost for $SO_2$ control. The NSPS does not prohibit bypassing a portion of the flue gas for flue gas reheat, provided the overall emission limitation is met.

Another requirement of the 1979 NSPS specifies continuous compliance via continuous emission monitors (CEM) to determine compliance with the standard. The standard requires that the $SO_2$ removal be met on a 30-day rolling average. Two data points per hour, 18 hours per day, and 22 days per 30-day period are the minimum data requirements of the standard. Results obtained from continuous emission monitoring of several plants conducted by the EPA are presented in Section 4.

The influences of the 1979 NSPS are visible in the design of commercial systems and in the cost estimates made by various groups involved with FGD. Nearly all new systems have spare module capacity. The spray tower absorber design is now the most popular, due to its internal simplicity and resulting lower maintenance requirements. Bypass ducts are being installed and costed in the process estimates.

The performance of FGD systems has been improved over the years to the point that the process economics of the different technologies are often the deciding factor in specific applications. Consequently, limestone systems, until recently the lowest cost systems, are also the most prevalent. Spray drying systems, because of their even lower costs, could eventually become the most popular technology.

## COMMERCIAL FGD PROCESS DESCRIPTIONS

Eight types of FGD systems are, or have been commercially operated. These systems are: limestone, lime, spray drying, alkaline ash, sodium carbonate, Wellman-Lord, dual alkali, and magnesium oxide. Both the Citrate and Aqueous Carbonate processes are involved in demonstration programs and are discussed with the developing processes at the end of this section. Information about each process is presented in a similar manner. An introduction discusses the status of each technology with regard to commercial applications. A chemistry and process description is presented including performance characteristics and waste or byproduct disposal options, followed by a discussion of the impacts of the 1979 NSPS on the technology. Recent innovations and areas of continuing development are then presented. The final section discusses process economics.

Table 13 presents the power generating stations identified in the PEDCo data base as having selected an FGD technology. The stations are presented alphabetically and show the operating utility, unit size, operational status, and regulatory status. It should be noted that the unit size is expressed as the equivalent flue gas capacity and not always the gross MW. Some units are legally permitted to operate with only partial scrubbing of the flue gas. All further tables in this report only identify the FGD unit and not the operating utility.

### Limestone

The limestone FGD system is the most prevalent process used. The total megawatts of power controlled by limestone systems is over twice the mega-

TABLE 13. ALPHABETICAL LIST OF UTILITY FGD SYSTEMS

| Unit | Utility | FGD Process | MW | Status[a] | Reg Class[b] |
|---|---|---|---|---|---|
| Antelope Valley 1 | Basin Electric Power | Spray Drying | 440 | 2 | D |
| Antelope Valley 2 | Basin Electric Power | Spray Drying | 440 | 2 | C |
| Apache 2 | Arizona Electric Power | Limestone | 98 | 1 | D |
| Apache 3 | Arizona Electric Power | Limestone | 98 | 1 | D |
| Arkansas Lignite 5 | Middle South Utilities | Limestone | 890 | 4 | A |
| Arkansas Lignite 6 | Middle South Utilities | Limestone | 890 | 4 | A |
| Big Bend 4 | Tampa Electric | Limestone | 475 | 3 | A |
| Bonanza 1 | Deseret Gen. and Trans. | Limestone | 410 |  | A |
| Bonanza 2 | Deseret Gen. and Trans. | Limestone | 410 | 5 | A |
| Clay Boswell 4 | Minnesota Power & Light | Alkaline Ash | 475 | 1 | A |
| Jim Bridger 1 | Pacific Power & Light | Sodium Carbonate | 550 | 6 | A |
| Jim Bridger 2 | Pacific Power & Light | Sodium Carbonate | 550 | 6 | A |
| Jim Bridger 2A | Pacific Power & Light | Spray Drying | 100 | 2 | A |
| Jim Bridger 3 | Pacific Power & Light | Sodium Carbonate | 550 | 1 | D |
| Jim Bridger 4 | Pacific Power & Light | Sodium Carbonate | 550 | 1 | D |
| A.B. Brown 1 | Southern Indiana Gas & Elec. | Dual Alkali | 265 | 1 | B |
| A.B. Brown 2 | Southern Indiana Gas & Elec. | Dual Alkali | 265 | 5 | A |
| Cane Run 4 | Louisville Gas & Electric | Lime | 188 | 1 | D |
| Cane Run 5 | Louisville Gas & Electric | Lime | 200 | 1 | D |
| Cane Run 6 | Louisville Gas & Electric | Dual Alkali | 299 | 1 | D |
| Cholla 1 | Arizona Public Service | Limestone | 119 | 1 | C |
| Cholla 2 | Arizona Public Service | Limestone | 264 | 1 | C |
| Cholla 4 | Arizona Public Service | Limestone | 126 | 1 | C |
| Coal Creek 1 | Cooperative Power Assoc. | Alkaline Ash | 327 | 1 | B |
| Coal Creek 2 | Cooperative Power Assoc. | Alkaline Ash | 327 | 1 | B |
| Colstrip 1 | Montana Power | Alkaline Ash | 360 | 1 | B |
| Colstrip 2 | Montana Power | Alkaline Ash | 360 | 1 | B |
| Colstrip 3 | Montana Power | Alkaline Ash | 700 | 2 | C |
| Colstrip 4 | Montana Power | Alkaline Ash | 700 | 2 | C |
| Conesville 5 | Columbus & So. Ohio Elec. | Lime | 411 | 1 | D |
| Conesville 6 | Columbus & So. Ohio Elec. | Lime | 411 | 1 | D |
| Coronado 1 | Salt River Project | Limestone | 280 | 1 | D |
| Coronado 2 | Salt River Project | Limestone | 280 | 1 | D |
| Coronado 3 | Salt River Project | Limestone | 400 | 6 | D |
| Coyote 1 | Montana-Dakota Utilities | Spray Drying | 440 | 1 | B |
| Craig 1 | Colorado Ute Electric | Limestone | 410 | 1 | D |
| Craig 2 | Colorado Ute Electric | Limestone | 410 | 1 | D |
| Craig 3 | Colorado Ute Electric | Spray Drying | 447 | 2 | D |
| Creston Coal 1 | Washington Water & Power | Limestone | 570 | 6 | A |
| Creston Coal 2 | Washington Water & Power | Limestone | 570 | 6 | C |
| Creston Coal 3 | Washington Water & Power | Limestone | 570 | 6 | C |
| Creston Coal 4 | Washington Water & Power | Limestone | 570 | 6 | C |
| Cromby | Philadelphia Electric | Magnesium Oxide | 150 | 2 | D |
| Cross 1 | So. Carolina Public Service | Limestone | 500 | 3 | A |
| Cross 2 | So. Carolina Public Service | Limestone | 500 | 2 | A |

(Continued)

TABLE 13. ALPHABETICAL LIST OF UTILITY FGD SYSTEMS (Continued)

| Unit | Utility | FGD Process | MW | Status[a] | Reg Class[b] |
|---|---|---|---|---|---|
| Dallman 3 | Springfield Water, Light and Power | Limestone | 185 | 1 | B |
| Delaware City 1 | Delmarva Power & Light | Wellman-Lord | 60 | 1 | E |
| Delaware City 2 | Delmarva Power & Light | Wellman-Lord | 60 | 1 | E |
| Delaware City 3 | Delmarva Power & Light | Wellman-Lord | 60 | 1 | E |
| Dolet Hills 1 | Southwestern Elec. Power | Limestone | 720 | 3 | A |
| Dolet Hills 2 | Southwestern Elec. Power | Limestone | 720 | 6 | A |
| Duck Creek 1 | Central Illinois Light | Limestone | 416 | 1 | B |
| Duck Creek 2 | Central Illinois Light | Limestone | 450 | 6 | A |
| East Bend 2 | Cinncinati Gas & Electric | Lime | 650 | 1 | B |
| Eddystone 1 | Philadelphia Electric | Magnesium Oxide | 240 | 2 | D |
| Eddystone 2 | Philadelphia Electric | Magnesium Oxide | 334 | 2 | D |
| Elrama 1-4 | Duquesne Light | Lime | 510 | 1 | D |
| Fayette 3 | Lower Colorado River Auth. | Limestone | 435 | 6 | A |
| Four Corners 1 | Arizona Public Service | Alkaline Ash | 175 | 1 | C |
| Four Corners 2 | Arizona Public Service | Alkaline Ash | 175 | 1 | C |
| Four Corners 3 | Arizona Public Service | Alkaline Ash | 229 | 1 | C |
| Four Corners 4 | Arizona Public Service | Lime | 755 | 2 | C |
| Four Corners 5 | Arizona Public Service | Lime | 755 | 2 | C |
| Reid Gardner 1 | Nevada Power | Sodium Carbonate | 125 | 1 | B |
| Reid Gardner 2 | Nevada Power | Sodium Carbonate | 125 | 1 | B |
| Reid Gardner 3 | Nevada Power | Sodium Carbonate | 125 | 1 | B |
| Reid Gardner 4 | Nevada Power | Sodium Carbonate | 250 | 4 | A |
| Gibbons Creek 1 | Texas Municipal Power Agcy. | Limestone | 400 | 2 | A |
| Gibson 5 | Public Service of Indiana | Limestone | 650 | 2 | B |
| Grand River 2 | Grand River Dam Authority | Spray Drying | 520 | 3 | A |
| Green 1 | Big Rivers Electric | Lime | 242 | 1 | B |
| Green 2 | Big Rivers Electric | Lime | 242 | 1 | B |
| Green River 1-3 | Kentucky Utilities | Lime | 64 | 1 | E |
| Guthrie Co. 1 | Iowa Electric Light & Power | Limestone | 720 | 4 | A |
| Hancock 1 | Kentucky Utilities | Limestone | 650 | 4 | A |
| Hancock 2 | Kentucky Utilities | Limestone | 650 | 6 | A |
| Hawthorne 3 | Kansas City Power & Light | Lime | 90 | 1 | D |
| Hawthorne 4 | Kansas City Power & Light | Lime | 90 | 1 | D |
| Thomas Hill 3 | Associated Electric | Limestone | 670 | 2 | A |
| Holcomb 1 | Sunflower Electric | Spray Drying | 347 | 2 | A |
| Hunter 1 | Utah Power & Light | Lime | 360 | 1 | B |
| Hunter 2 | Utah Power & Light | Lime | 360 | 1 | B |
| Hunter 3 | Utah Power & Light | Limestone | 400 | 2 | A |
| Hunter 4 | Utah Power & Light | Limestone | 400 | 2 | A |
| Huntington 1 | Utah Power & Light | Lime | 366 | 1 | B |
| Charles R. Huntley 66 | Niagra Mohawk Power | Aqueous Carbonate | 100 | 2 | E |

(Continued)

TABLE 13. ALPHABETICAL LIST OF UTILITY FGD SYSTEMS (Continued)

| Unit | Utility | FGD Process | MW | Status[a] | Reg Class[b] |
|---|---|---|---|---|---|
| Intermountain 1 | Los Angeles Dept. of Water and Power | Limestone | 820 | 6 | A |
| Intermountain 2 | Los Angeles Dept. of Water and Power | Limestone | 820 | 6 | A |
| Intermountain 3 | Los Angeles Dept. of Water and Power | Limestone | 820 | 6 | A |
| Intermountain 4 | Los Angeles Dept. of Water and Power | Limestone | 820 | 6 | A |
| Jeffery 1 | Kansas Power & Light | Limestone | 540 | 1 | D |
| Jeffery 2 | Kansas Power & Light | Limestone | 490 | 1 | D |
| LaCygne 1 | Kansas City Power & Light | Limestone | 874 | 1 | E |
| Lawrence 4 | Kansas Power & Light | Limestone | 125 | 1 | D |
| Lawrence 5 | Kansas Power & Light | Limestone | 420 | 1 | D |
| Laramie River 1 | Basin Electric Power | Limestone | 570 | 1 | C |
| Laramie River 2 | Basin Electric Power | Limestone | 570 | 1 | C |
| Laramie River 3 | Basin Electric Power | Spray Drying | 570 | 2 | C |
| Limestone 1 | Houston Lighting & Power | Limestone | 750 | 3 | A |
| Limestone 2 | Houston Lighting & Power | Limestone | 750 | 3 | A |
| Bruce Mansfield 1 | Pennsylvania Power | Lime | 917 | 1 | D |
| Bruce Mansfield 2 | Pennsylvania Power | Lime | 917 | 1 | D |
| Bruce Mansfield 3 | Pennsylvania Power | Lime | 917 | 1 | D |
| Marion 4 | Southern Illinois Power | Limestone | 173 | 1 | B |
| Martin Lake 1 | Texas Utilities | Limestone | 595 | 1 | B |
| Martin Lake 2 | Texas Utilities | Limestone | 595 | 1 | B |
| Martin Lake 3 | Texas Utilities | Limestone | 595 | 1 | B |
| Martin Lake 4 | Texas Utilities | Limestone | 750 | 3 | A |
| McIntosh 3 | Lakeland Utilities | Limestone | 364 | 2 | A |
| Merom 1 | Hoosier Energy | Limestone | 441 | 2 | B |
| Merom 2 | Hoosier Energy | Limestone | 441 | 1 | B |
| Metro Coal | Northern States Power | Spray Drying | 200 | 6 | A |
| Mill Creek 1 | Louisville Gas & Electric | Lime | 358 | 1 | E |
| Mill Creek 2 | Louisville Gas & Electric | Lime | 350 | 1 | E |
| Mill Creek 3 | Louisville Gas & Electric | Lime | 427 | 1 | D |
| Mill Creek 4 | Louisville Gas & Electric | Lime | 495 | 2 | B |
| Dean H. Mitchell 11 | No. Indiana Public Service | Wellman-Lord | 115 | 1 | E |
| Mitchell 33 | West Pennsylvania Power | Lime | 300 | 2 | C |
| Montezuma 1 | Pacific Gas & Electric | Limestone | 800 | 6 | C |
| Montezuma 2 | Pacific Gas & Electric | Limestone | 800 | 6 | C |
| Monticello 3 | Texas Utilities | Limestone | 800 | 1 | B |
| R.D. Morrow 1 | So. Mississippi Elec. Pwr. | Limestone | 124 | 1 | B |
| R.D. Morrow 2 | So. Mississippi Elec. Pwr. | Limestone | 124 | 1 | B |
| Muscatine 9 | Muscatine Power & Water | Limestone | 166 | 2 | B |

(Continued)

TABLE 13. ALPHABETICAL LIST OF UTILITY FGD SYSTEMS (Continued)

| Unit | Utility | FGD Process | MW | Status[a] | Reg Class[b] |
|---|---|---|---|---|---|
| Naughton 3 | Utah Power & Light | Sodium Carbonate | 330 | 1 | C |
| Newton 1 | Central Illinois Public Service | Dual Alkali | 617 | 1 | B |
| Oxbow 1 | Cajun Electric Power | Spray Drying | 540 | 6 | A |
| Paddy's Run 6 | Louisville Gas & Electric | Lime | 72 | 1 | E |
| Paradise 1 | Tennessee Valley Authority | Limestone | 704 | 2 | E |
| Paradise 2 | Tennessee Valley Authority | Limestone | 704 | 2 | E |
| W.A. Parish 8 | Houston Lighting & Power | Limestone | 492 | 2 | A |
| Patriot 1 | Indianapolis Power & Light | Limestone | 650 | 6 | C |
| Patriot 2 | Indianapolis Power & Light | Limestone | 650 | 6 | C |
| Patriot 3 | Indianapolis Power & Light | Limestone | 650 | 6 | C |
| Petersburg 3 | Indianapolis Power & Light | Limestone | 532 | 1 | B |
| Petersburg 4 | Indianapolis Power & Light | Limestone | 530 | 2 | B |
| Phillips 1-6 | Duquesne Light | Lime | 408 | 1 | D |
| Henry W. Pirkey 1 | Southwestern Electric Power | Limestone | 720 | 3 | B |
| Plains Escalante 1 | Plains Escalante G & T | Limestone | 233 | 2 | B |
| Pleasants 1 | Monongahela Power | Lime | 668 | 1 | B |
| Pleasants 2 | Monongahela Power | Lime | 668 | 1 | B |
| Powerton 51 | Commonwealth Edison | Limestone | 450 | 1 | E |
| Project 1 | Michigan South Central Power Agency | Limestone | 55 | 2 | A |
| Rawhide 1 | Platte River Power Authority | Spray Drying | 279 | 3 | C |
| Riverside 6-7 | Northern States Power | Srpay Drying | 110 | 1 | E |
| St. Johns 1 | Jacksonville Elec. Authority | Limestone | 600 | 5 | A |
| St. Johns 2 | Jacksonville Elec. Authority | Limestone | 600 | 5 | A |
| Sandow 4 | Texas Power & Light | Limestone | 382 | 1 | B |
| San Juan 1 | Public Service of New Mexico | Wellman-Lord | 361 | 1 | C |
| San Juan 2 | Public Service of New Mexico | Wellman-Lord | 350 | 1 | C |
| San Juan 3 | Public Service of New Mexico | Wellman-Lord | 534 | 1 | C |
| San Juan 4 | Public Service of New Mexico | Wellman-Lord | 534 | 2 | C |
| San Miguel 1 | San Miguel Electric | Limestone | 400 | 1 | B |
| Schahfer 17 | No. Indiana Public Service | Dual Alkali | 421 | 2 | A |
| Schahfer 18 | No. Indiana Public Service | Dual Alkali | 421 | 3 | A |
| Seminole 1 | Seminole Electric | Limestone | 620 | 2 | A |
| Seminole 2 | Seminole Electric | Limestone | 620 | 3 | A |
| Shawnee 10A | Tennessee Valley Authority | Limestone | 10 | 1 | E |
| Shawnee 10B | Tennessee Valley Authority | Limestone | 10 | 1 | E |
| Sherburne 1 | Northern States Power | Alkaline Ash | 740 | 1 | D |
| Sherburne 2 | Northern States Power | Alkaline Ash | 740 | 1 | D |
| Sherburne 3 | Northern States Power | Spray Drying | 810 | 6 | C |
| Shiras 3 | Marquette Board of Light and Power | Spray Drying | 44 | 2 | A |
| Sikeston 1 | Sikeston Brd. of Mun. Util. | Limestone | 235 | 1 | B |

(Continued)

TABLE 13. ALPHABETICAL LIST OF UTILITY FGD SYSTEMS (Continued)

| Unit | Utility | FGD Process | MW | Status[a] | Reg Class[b] |
|---|---|---|---|---|---|
| J.B. Sims 3 | Grand Haven Brd. of Light and Power | Lime | 65 | 2 | A |
| J.K. Smith 1 | East Kentucky Power | Lime | 650 | 3 | A |
| Somerset 1 | New York State Electric and Gas | Limestone | 625 | 4 | E |
| Southwest 1 | Springfield City Utilities | Limestone | 194 | 1 | B |
| Spingerville 1 | Tuscon Electric Power | Spray Drying | 370 | 3 | B |
| Spingerville 2 | Tuscon Electric Power | Spray Drying | 370 | 3 | B |
| Spurlock 2 | East Kentucky Power | Lime | 500 | 2 | B |
| Stanton 10 | United Power Association | Spray Drying | 50 | 1 | A |
| Tombigbee 2 | Alabama Electric | Limestone | 179 | 1 | B |
| Tombigbee 3 | Alabama Electric | Limestone | 179 | 1 | B |
| Twin Oaks 1 | Texas Power & Light | Limestone | 750 | 3 | A |
| Twin Oaks 2 | Texas Power & Light | Limestone | 750 | 3 | A |
| Unassigned 1 | Middle South Utilities | Limestone | 890 | 4 | A |
| Unassigned 2 | Middle South Utilities | Limestone | 890 | 4 | A |
| Valmy 2 | Sierra Pacific Power | Spray Drying | 276 | 6 | A |
| G.F. Weston | St. Joseph Zinc | Citrate | 60 | 1 | B |
| Widows Creek 7 | Tennessee Valley Authority | Limestone | 575 | 1 | E |
| Widows Creek 8 | Tennessee Valley Authority | Limestone | 550 | 1 | E |
| D.B. Wilson 1 | Big Rivers Electric | Limestone | 440 | 2 | A |
| D.B. Wilson 2 | Big Rivers Electric | Limestone | 440 | 3 | A |
| Wilton 1 | Middle South Utilities | Limestone | 890 | 4 | A |
| Wilton 2 | Middle South Utilities | Limestone | 890 | 4 | A |
| Winyah 2 | So. Carolina Pub. Service | Limestone | 140 | 1 | B |
| Winyah 3 | So. Carolina Pub. Service | Limestone | 280 | 1 | B |
| Winyah 4 | So. Carolina Pub. Service | Limestone | 280 | 1 | A |
| Milton R. Young 2 | Minnkota Power | Alkaline Ash | 185 | 1 | D |

[a] Status: 1 - Operational; 2 - Under construction; 3 - Contract awarded; 4 - Letter of intent signed; 5 - Requesting/evaluating bids; 6 - Considering FGD.

[b] Regulatory Classification: A - 1979 NSPS; B - 1971 NSPS; C - Standard more stringent than 1979 NSPS; D - Standard more stringent than 1971 NSPS but less than 1979 NSPS; E - Standard equal to or less than 1971 NSPS.

Source: Reference 2.

watts controlled by lime systems, the second most popular process. There are three major reasons for this dominance. First, limestone and chemically similar lime systems have been in use in the U.S. for over 10 years. Consequently, a large amount of experience has been obtained resulting in improvements in operations and control. Secondly, the reagent cost for limestone is much lower than lime or sodium reagents. This reduces the annualized costs of operation. Third, by varying the system operating parameters, the process can treat a variety of coals fairly efficiently.

Table 14 presents the operational and regulatory status of limestone FGD systems as of January 1983. The applicable $SO_2$ standards determine the degree of $SO_2$ removal needed for each unit. The systems regulated by the 1979 NSPS are primarily in the construction and planning stages. The majority of the units regulated by the 1971 NSPS are now in operation. The other regulatory classifications are due to state requirements. Several Western states have required emissions to be much lower than the Federal NSPS (0.2-0.4 lb $SO_2/10^6$ Btu). Many of the systems which do not meet the 1971 NSPS are for large, high sulfur, retrofit systems.

Process Description—

A generalized flow diagram of a limestone FGD system is presented in Figure 4. Flue gas and slurry are contacted in the absorber, where $SO_2$ is removed. The slurry flows to the reaction tank where limestone is dissolved and calcium sulfite and sulfate are precipitated as shown in the following overall reactions.

$$SO_2(g) + CaCO_3(1) + 1/2\ H_2O \rightleftarrows CaSO_3 \cdot 1/2\ H_2O(s) + CO_2(g)$$

$$SO_3^{=} + 1/2O_2(g) \rightarrow SO_4^{=}$$

$$SO_4^{=} + Ca^{++} + 2H_2O \rightleftarrows CaSO_4 \cdot 2H_2O(s)$$

A bleed stream of slurry is often concentrated by a thickener, vacuum filter, or centrifuge before it is disposed of. Sludge ponds and landfills

TABLE 14. OPERATIONAL AND REGULATORY STATUS OF LIMESTONE FGD SYSTEMS AS OF JANUARY 1983

| Applicable $SO_2$ Standard | MW Capacity (number of units) | | | | |
|---|---|---|---|---|---|
| | Operating | In Construction | Contracted | Planned | Total |
| 1979 NSPS | 280 (1) | 4,751 (11) | 6,505 (10) | 15,925 (23) | 27,461 (45) |
| 1971 NSPS | 6,569 (19) | 2,020 (5) | 720 (1) | – | 9,309 (25) |
| >1979 NSPS | 1,649 (5) | – | – | 5,590 (9) | 7,239 (14) |
| >1971 NSPS, <1979 NSPS | 3,151 (10) | – | – | – | 3,151 (10) |
| ≤ 1971 NSPS | 2,469 (6) | 1,408 (2) | – | 625 (1) | 4,502 (9) |
| TOTAL | 14,118 (41) | 8,179 (18) | 7,225 (11) | 21,810 (33) | 51,662 (103) |

Source: Reference 2.

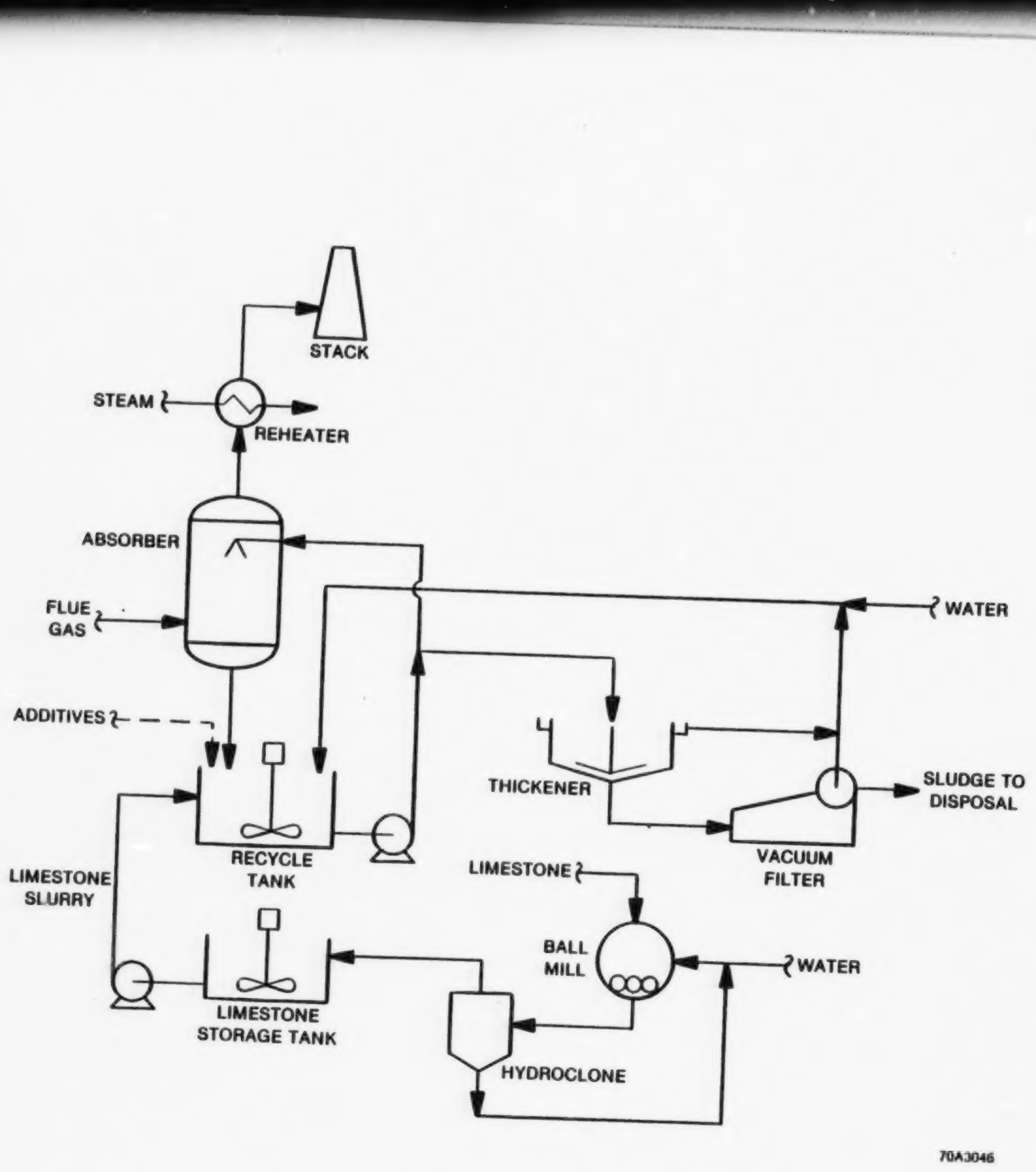

STACK
STEAM
REHEATER
ABSORBER
FLUE GAS
ADDITIVES
RECYCLE TANK
LIMESTONE SLURRY
LIMESTONE STORAGE TANK
WATER
THICKENER
SLUDGE TO DISPOSAL
VACUUM FILTER
LIMESTONE
BALL MILL
WATER
HYDROCLONE
70A3046

Figure 4. Limestone Process Flow Diagram

are the two most prevalent disposal options. The limestone is prepared for use typically in a ball mill where rock size particles are crushed and form a slurry. Limestone addition is regulated to maintain the desired pH in the reaction tank. Water is added to compensate for sludge and evaporative losses.

The chemistry of limestone and all other FGD systems is fairly complex, involving mass transfer between gas, liquid, and solid phases. Many other reports discuss in detail the reactions occuring in a limestone system (23, 24, 25). One important consideration in the design of all calcium based systems is the relative insolubility of most calcium compounds. This factor requires large amounts of liquid slurry per volume of gas (high L/G ratios) to obtain effective $SO_2$ absorption and neutralization. High L/G rates and various additives can be used to design systems capable of 90 percent $SO_2$ removal from high sulfur coals. This topic is discussed further in the recent innovations section. The low solubility has also caused scaling problems when the liquid becomes supersaturated with respect to calcium sulfate. If insufficient seed crystals are available, scaling can occur on the system internals. This can raise the gas-side pressure drop and generally requires manual cleaning to remove the scale.

The major drawback of all throwaway systems is the large volumes of solid waste generated. Conventional limestone systems produce a sludge composed primarily of calcium sulfite, which because of crystalline properties, poses handling problems. This sludge reliquifies upon application of stress and does not dewater well. Consequently, ponds are the normal method of ultimate disposal, however, further problems can arise due to process liquor infiltrating the ground water. Plastic or clay liners are usually required to prevent this type of contamination. Currently, FGD wastes are classified as non-hazardous according to RCRA regulations, pending the outcome of an EPA study examining these wastes (26). Another recent innovation, forced oxidation, is being employed at many new systems to improve the sludge handling properties by producing gypsum.

Commercial Designs--

Table 15 presents several design features of the limestone FGD systems identified in the PEDCo survey. The various systems have been grouped according to the five regulatory classifications for $SO_2$ emissions discussed previously to illustrate process design and operating trends associated with each regulatory class. As can be seen by examining this table, there are several areas in which the new NSPS has impacted the system design.

The first question which arises when considering the impact of the revised NSPS involves the percent sulfur in the coal selected for new units. The average sulfur content of limestone units regulated by the 1979 NSPS is 1.47 percent. Units controlled by the 1971 standard average 2.08 percent sulfur. This would appear to indicate that the new NSPS is increasing the use of low sulfur coal relative to high sulfur coal. However, under the 1971 NSPS, low sulfur coal could be used instead of scrubbing, provided the emission rate was less than 1.2 pounds of $SO_2$ per million Btu of heat input. Of the coalfired boilers that came on-line between 1976 and 1984, 90 units used FGD systems while 102 units did not (22). A large number of the units built without FGD systems used low sulfur coal to achieve compliance. Including the sulfur content of these units with the average sulfur content of the units with FGD systems controlled by the 1971 NSPS would bring the average sulfur closer to that found with the 1979 NSPS. Fifteen of the 39 units in Table 15, regulated by the new standard, burn 1.5 percent sulfur or higher coal. This is approximately the same ratio as previously given for the 1971 NSPS units, 90 FGD systems out of 192 units. Figure 5 shows the sulfur content of all units with various FGD systems controlled by the 1971 and 1979 NSPS. This figure supports the previous contention that new systems are not selectively using low sulfur coal to lower FGD system costs. The ratio of high to low sulfur units would be lower for the 1971 NSPS data if units which did not require FGD systems were included.

## TABLE 15. LIMESTONE FGD SYSTEM DESIGN

| Plant | Startup | Percent Sulfur | L/G (gal/1000 acf) | Absorber Removal (%) | Unit Removal (%) | Absorber Type[a] | Vendor[b] |
|---|---|---|---|---|---|---|---|
| | | | 1979 NSPS | | | | |
| Thomas Hill 3 | 01/82 | 4.8 | – | 92 | 92 | ST | K |
| D.B. Wilson 1 | 07/84 | – | – | – | – | ST | K |
| D.B. Wilson 2 | – | – | – | – | – | ST | K |
| Duck Creek 2 | 01/87 | 3.3 | – | – | – | – | – |
| Bonanza 1 | 09/84 | 0.5 | – | – | 95 | – | CE |
| Bonanza 2 | 00/88 | 0.5 | – | – | 95 | – | – |
| Limestone 1 | 12/86 | 1.1 | 122 | 90 | 90 | ST | CE |
| Limestone 2 | 12/87 | 1.1 | 122 | 90 | 90 | ST | CE |
| W.A. Parish 8 | 11/82 | 0.6 | 42 | 85 | 70 | ST | GE |
| Guthrie Co. 1 | 11/87 | 0.4 | – | – | – | – | CE |
| St. Johns 1 | 12/85 | 2.5 | – | – | – | – | – |
| St. Johns 2 | 06/87 | 2.5 | – | – | – | – | – |
| Hancock 1 | 04/89 | 3.5 | – | – | – | – | BW |
| Hancock 2 | 04/94 | 3.5 | – | – | – | – | – |
| McIntosh 3 | 01/82 | 2.6 | 80 | 87 | 87 | ST | BW |
| Fayette 3 | 00/88 | 1.7 | – | – | 90 | – | – |
| Project 1 | 06/82 | 2.3 | – | 90 | 90 | ST | BW |
| Arkansas 1 | 00/90 | 0.5 | – | – | – | – | CE |
| Arkansas 2 | 00/92 | 0.5 | – | – | – | – | CE |
| Unassigned 1 | 00/89 | 0.5 | – | – | – | – | CE |
| Unassigned 2 | 00/93 | 0.5 | – | – | – | – | CE |
| Wilton 1 | 00/88 | 0.5 | – | – | – | – | CE |
| Wilton 2 | 00/91 | 0.5 | – | – | – | – | CE |
| Coronado 3 | 00/89 | 0.6 | – | – | – | – | – |
| Seminole 1 | 03/83 | 2.8 | 53 | 86 | 86 | ST | P |
| Seminole 2 | 01/85 | 2.8 | 53 | 86 | 86 | ST | P |
| Cross 1 | 05/88 | 1.8 | – | – | 86 | ST | P |
| Cross 2 | 11/83 | 1.8 | – | – | 86 | ST | P |
| Winyah 4 | 07/81 | 1.7 | – | 80 | 80 | ST | AAF |
| Dolet Hills 1 | 03/86 | 0.7 | – | – | – | – | UOP |
| Dolet Hills 2 | 00/88 | 0.7 | – | – | – | – | – |
| Big Bend 4 | 03/85 | 2.4 | 92 | 90 | 90 | G | RC |
| Gibbons Creek 1 | 10/82 | 1.1 | – | – | 87 | ST | CE |
| Twin Oaks 1 | 08/85 | 0.7 | – | – | 70 | – | GE |
| Twin Oaks 2 | 08/87 | 0.7 | – | – | 70 | – | GE |
| Martin Lake 4 | 00/89 | 0.9 | – | – | 75 | G | RC |
| Hunter 3 | 06/83 | 0.6 | – | – | – | ST | GE |
| Hunter 4 | 06/85 | 0.6 | – | – | – | ST | GE |
| Creston Coal 1 | 00/87 | – | – | – | – | – | – |

(Continued)

TABLE 15. LIMESTONE FGD SYSTEM DESIGN (Continued)

| Plant | Startup | Percent Sulfur | L/G (gal/1000 acf) | Absorber Removal (%) | Unit Removal (%) | Absorber Type[a] | Vendor[b] |
|---|---|---|---|---|---|---|---|
| | | | 1971 NSPS | | | | |
| Tombigbee 2 | 09/78 | 1.1 | 74 | 85 | 60 | ST | P |
| Tombigbee 3 | 06/79 | 1.1 | 74 | 85 | 60 | ST | P |
| Duck Creek 1 | 07/76 | 3.7 | 53 | 85 | 85 | G | E-RS |
| Merom 1 | 05/82 | 3.5 | – | 90 | – | G | M |
| Merom 2 | 12/81 | 3.5 | – | 90 | – | G | M |
| Petersburg 3 | 12/77 | 3.3 | 63 | 85 | 85 | P | UOP |
| Petersburg 4 | 10/84 | 3.5 | 63 | 85 | 85 | G | RC |
| Muscatine 9 | 09/82 | 3.2 | – | 94 | – | G | RC |
| Plains Escalante 1 | 12/83 | 0.8 | – | – | – | – | CE |
| Gibson 5 | 08/82 | 3.3 | – | – | – | ST | K |
| San Miguel | 08/81 | 1.7 | 50 | 86 | 86 | T | BW |
| Sikeston 1 | 06/81 | 2.8 | 76 | 80 | 80 | V | BW |
| Winyah 2 | 07/77 | 1.1 | 58[c] | 90 | 45 | V-T | BW |
| Winyah 3 | 05/80 | 1.1 | – | 90 | 90 | ST | BW |
| R.D. Morrow 1 | 08/78 | 1.3 | 56 | 85 | 53 | G | E-RS |
| R.D. Morrow 2 | 06/79 | 1.3 | 56 | 85 | 53 | G | E-RS |
| Marion 4 | 05/79 | 3.8 | 62[c] | 89 | 89 | V-ST | BW |
| H.W. Pirkey 1 | 12/84 | 0.8 | – | – | – | ST | UOP |
| Southwest 1 | 04/77 | 3.5 | 50 | 87 | 80 | P | UOP |
| Dallman 3 | 12/80 | 3.3 | 78[c] | 95 | 95 | G | RC |
| Sandow 4 | 12/80 | 1.6 | – | – | 75 | ST | CE |
| Martin Lake 1 | 04/77 | 0.9 | – | 95 | 71 | G | RC |
| Martin Lake 2 | 05/78 | 0.9 | – | 95 | 71 | G | RC |
| Martin Lake 3 | 02/79 | 0.9 | – | 95 | 71 | G | RC |
| Monticello 3 | 05/78 | 1.5 | 70 | 74 | 74 | ST | GE |
| | | | Stricter Standard Than 1979 NSPS | | | | |
| Cholla 1 | 10/73 | 0.5 | 49 | 92 | 58 | G | RC |
| Cholla 2 | 04/78 | 0.5 | 49 | 75 | 75 | G | RC |
| Cholla 4 | 03/81 | 0.5 | – | 95 | 95 | G | RC |
| Laramie River 1 | 07/80 | 0.8 | 77 | 90 | 90 | G | RC |
| Patriot 1 | 00/87 | 3.5 | – | – | – | – | – |
| Patriot 2 | – | 3.5 | – | – | – | – | – |
| Patriot 3 | – | 3.5 | – | – | – | – | – |
| Montezuma 1 | 06/93 | 0.8 | – | – | – | – | – |
| Montezuma 2 | 06/94 | 0.8 | – | – | – | – | – |
| Creston Coal 2 | 00/87 | – | – | – | – | – | – |
| Creston Coal 3 | 00/87 | – | – | – | – | – | – |
| Creston Coal 4 | 00/87 | – | – | – | – | – | – |

(Continued)

TABLE 15. LIMESTONE FGD SYSTEM DESIGN (Continued)

| Plant | Startup | Percent Sulfur | L/G (gal/1000 acf) | Absorber Removal (%) | Unit Removal (%) | Absorber Type[a] | Vendor[b] |
|---|---|---|---|---|---|---|---|
| Stricter Standard Than 1971 NSPS but Less Than 1979 NSPS | | | | | | | |
| Apache 2 | 08/78 | 0.5 | 62 | 97 | 43 | G | RC |
| Apache 3 | 06/79 | 0.5 | 62 | 97 | 43 | G | RC |
| Craig 1 | 10/80 | 0.4 | 64 | 85 | 65 | ST | P |
| Craig 2 | 12/79 | 0.4 | 64 | 85 | 65 | ST | P |
| Jeffery 1 | 08/78 | 0.3 | 50 | 80 | 60 | ST | CE |
| Jeffery 2 | 01/80 | 0.3 | 50 | 80 | 60 | ST | CE |
| Lawrence 4 | 01/77 | 0.6 | 48[c] | 73 | 73 | V-ST | CE |
| Lawrence 5 | 04/78 | 0.6 | 35[c] | 52 | 52 | V-ST | CE |
| Coronado 1 | 11/79 | 0.5 | 23 | 82 | 66 | ST | K |
| Coronado 2 | 07/80 | 0.5 | 23 | 82 | 66 | ST | K |
| Standard Equal to or Less Than 1971 NSPS | | | | | | | |
| Powerton 51 | 04/80 | 3.5 | 64 | 75 | 75 | P | UOP |
| LaCygne 1 | 02/73 | 5.4 | 50[c] | 80 | 80 | V-T | BW |
| Somerset 1 | 06/84 | 2.2 | – | 90 | 90 | ST | P |
| Paradise 1 | 00/83 | 4.2 | – | 84 | 84 | V | GE |
| Paradise 2 | 00/83 | 4.2 | – | 84 | 84 | V | GE |
| Shawnee 10A | 04/72 | 2.9 | varies - test unit | | – | P | UOP |
| Shawnee 10B | 04/72 | 2.9 | varies - test unit | | – | V-ST | GE |
| Widows Creek 7 | 03/81 | 3.7 | – | 80 | 80 | V-ST | CE |
| Widows Creek 8 | 04/77 | 3.7 | 73[c] | 70 | 70 | V-P | TVA |

[a] Absorber Type: G - fixed grids; P - packed tower; ST - spray tower; T - tray tower; V - venturi.

[b] Vendors: AAF - American Air Filter; BW - Babcock and Wilcox; CE - Combustion Engineering;
E-RS - Environeering-Riley Stoker; GE - General Electric; K - Kellogg; M - Mitsubishi;
P - Peabody; RC - Research Cottrell; TVA - Tennessee Valley Authority;
UOP - Air Correction Division.

[c] Includes venturi or quencher L/G.

Source: Reference 2.

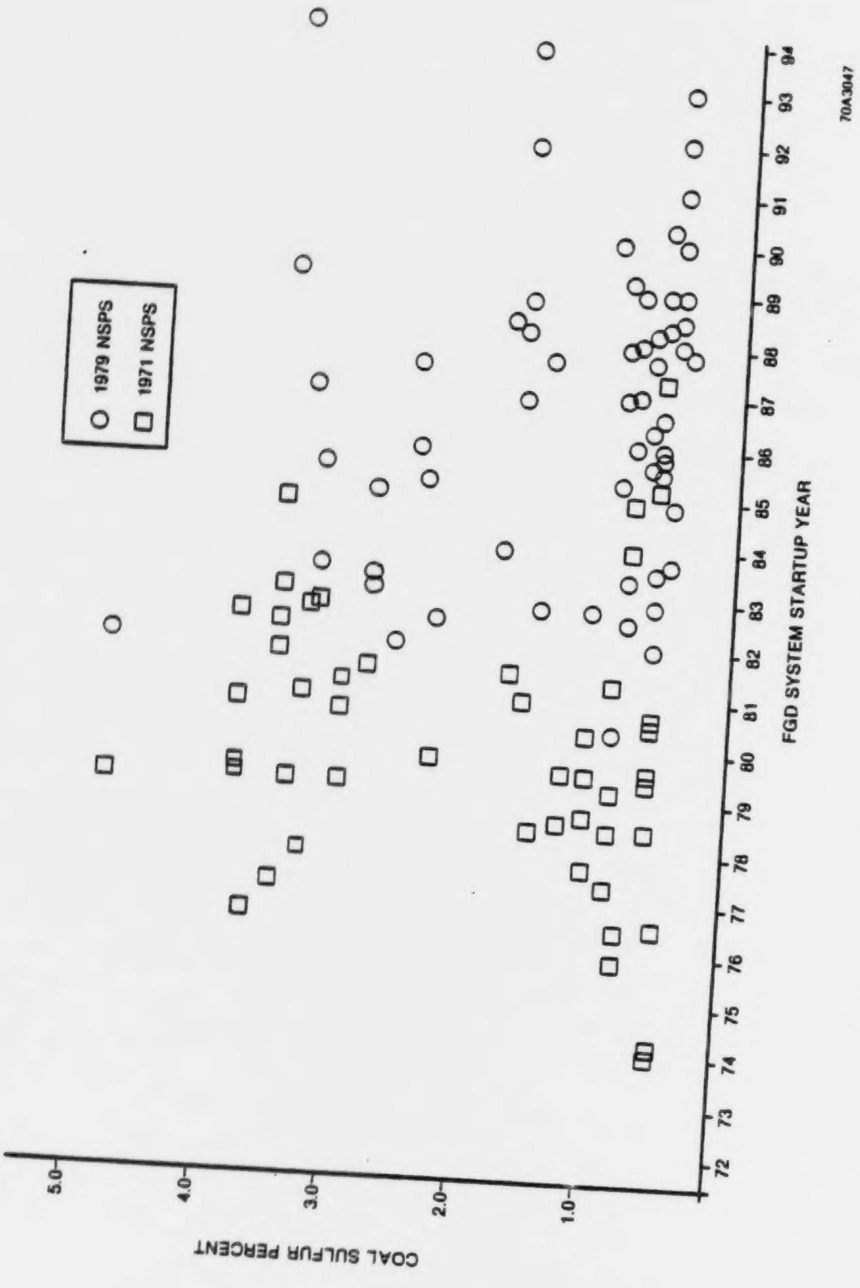


Figure 5. FGD System Sulfur Levels vs. Year of Startup

The liquid to gas ratio (L/G) of the various units show a slight increase with the stricter regulations; however, the variability of the reported values prevents this trend from being statistically significant. Likewise, the design absorber removal is slightly higher for the 1979 NSPS units, although nearly all of the absorbers are designed for 80 percent or greater removal. The overall unit removal efficiency does show a statistically significant increase from the 1971 NSPS average of 73.5 percent to the 1979 NSPS value of 85 percent overall removal.

The most obvious trend in commercial design is the type of absorbers used. The early lime and limestone absorbers were generally packed towers. Maintenance requirements often proved excessive due to plugging and scaling of the packing as well as accidental destruction due to temperature excursions. Simpler contactors such as grid towers and venturis have been used by many units, however, the open spray tower is the choice of the majority of new units. Table 16 presents the number and type of absorbers built by the various limestone FGD vendors in sequential order (based on start-up dates.) In all but one case, the undecided units will probably also be spray towers, based on the previous units installed by the vendor. If this trend continues, 43 of the 79 limestone units would be spray towers. Another 5 units are combination venturi-spray towers. Twenty units use grids with the remaining 11 units split among packed and tray towers and other venturi combinations.

The use of spray towers has been promoted by their higher reliability, compared to packed systems. The current NSPS allows bypass of unscrubbed gases only under limited circumstances, specifically, when the entire utility system is operating at full capacity and a spare module is in use. This requirement practically eliminates bypass of a malfunctioning absorber in all but the most critical of operating conditions. Consequently, reliability of the systems is a major factor in the design selection by the utility. This requirement has also resulted in the construction of a spare absorber module for almost all units larger than 125 MW. Most scrubber modules vary in size from 80 to 140 MW of equivalent gas capacity.

TABLE 16. SEQUENTIAL ORDER OF LIMESTONE ABSORBERS

| Vendor | Absorber Types[a] | Total |
|---|---|---|
| American Air Filter | ST | 1 |
| Babcock and Wilcox | V-T(2), V-ST, ST, V, T, ST(2), * | 9 |
| Combustion Engineering | V-ST, ST, V-ST, ST(2), V-ST, ST, *, ST(3), *(7) | 18 |
| Environeering-Riley Stoker | G(3) | 3 |
| General Electric | V-ST, ST(3), V(2), ST(3) | 9 |
| Kellogg | ST(6) | 6 |
| Mitsubishi | G(2) | 2 |
| Peabody | ST(9) | 9 |
| Research Cottrell | G(15) | 15 |
| TVA | V-P | 1 |
| UOP-Air Correction | P(4), ST, * | 6 |
|  |  | 79 |

[a] Absorber types: G - grids; P - packed; ST - spray tower; T - trays; V - venturi; * - unknown

Source: Reference 2.

Recent Innovations—

The limestone process has undergone refinements in design and operating conditions through the past several years to improve its performance and reliability. Some of the results of these research projects are currently being implemented at commercial installations. Three areas which offer process improvements are forced oxidation, $SO_2$ removal enhancement additives, and absorber design changes.

Forced Oxidation—Forced oxidation is a process modification in which air is sparged into a reaction tank to oxidize the calcium sulfite sludge to gypsum. Forced oxidation is used to improve the solids handling characteristics of the FGD solids. The reaction between $SO_2$ and limestone liberates $CO_2$ and forms a solid precipitate, calcium sulfite hemihydrate ($CaSO_3 \cdot 1/2H_2O$). Under the process conditions in an FGD system, the crystals produced have an irregular shape with a high void fraction. This shape causes the solids to settle slowly and dewater poorly. The high void fraction traps water in the sludge and can cause it to reliquefy upon agitation. These physical properties limit sludge disposal options primarily to ponding. The poor structural stability prevents landfilling or stacking the sludge.

Gypsum ($CaSO_4 \cdot 2H_2O$) has been produced as a by-product in the phosphate fertilizer industry for many years. For temporary and permanent disposal, the solids can be stacked to heights of over 100 feet because the gypsum crystal is a well defined rhombohedron which when compacted has a very low void fraction. A structurally stable solid mass is formed.

During the operation of early lime and limestone FGD systems, gypsum scaling on system internals was often a serious problem. Oxygen from the flue gas reacted with sulfite ions and formed sulfate ions. This "natural" oxidation is generally between 10 and 30 percent of the total $SO_2$ removed. It has been found that for oxidation of less than 15 percent, gypsum scaling does not occur (27). This is due to a coprecipitation mechanism in which sulfate ions are interspersed throughout the crystal lattice replacing sulfite ions. This coprecipitation mechanism can keep the scrubbing solution subsaturated with respect to gypsum. At oxidation levels about 15 percent, the coprecipitation mechanism is not capable of removing all of the sulfate ions from solution. Since very few gypsum seed crystals are present, crystal growth (scaling) on the system internals may occur to reduce the relative saturation.

Since the amount of oxidation that occurs is a function of the flue gas oxygen and $SO_2$ concentrations, the contactor design, and other relatively constant parameters, it is difficult to lower the oxidation fraction in operating systems if scaling is occurring. The alternative is to promote the oxidation to near completion. This provides sufficient gypsum seed crystal area to prevent scaling. Additionally, the sludge handling characteristics are greatly improved over unoxidized solids.

In general, low sulfur systems naturally operate in a high oxidation mode because of the high oxygen to $SO_2$ ratio in the flue gas. In most force oxidized systems, air is sparged into the reaction tank at a stoichiometry of 2-4 moles of atomic oxygen per mole of $SO_2$ removed. Other methods of oxidation have been examined by TVA at the Shawnee pilot scrubbers (28, 29). At least one commercial vendor uses a double loop system to produce gypsum. The second liquor loop, at a higher pH, is used to remove the bulk of the $SO_2$. Excess liquor is passed to the front loop where the pH drops and most of the oxidation takes place. This design also promotes good limestone utilization. This design will be used by Tampa Electric at the Big Bend 4 unit to produce saleable gypsum.

Additives—Both inorganic and organic additives have been used to improve the $SO_2$ removal efficiency of limestone systems. These additives are used because of the low solubility of calcium carbonate and calcium sulfite. This low solubility results in a low liquid phase alkalinity. With a low alkalinity concentration, it is necessary to contact the acidic flue gas with large slurry volumes. The liquid to gas ratio (L/G) for limestone systems is typically 60-120 gallons per thousand cubic feet of gas, depending on the $SO_2$ concentration and the design removal. Recirculating this large amount of slurry consumes a large portion of the system electrical demand. Maintenance on these very large pumps is often difficult and time consuming due to their size.

Several additives are being used commercially which increase the liquid phase alkalinity of limestone systems. Magnesium oxide is the most widely used additive. Adipic acid and waste dicarboxylic acids obtained during the production of adipic acid are also used commercially. The mechanism by which each additive enhances $SO_2$ removal is discussed below.

## Magnesium Oxide Additives

During the adsorption and neutralization of gaseous $SO_2$ and sulfurous acid ($H_2SO_3$), any species which is alkaline with respect to sulfurous acid can neutralize it, permitting continued $SO_2$ removal. The common species in a limestone system which are alkaline with respect to sulfurous acid are: carbonate, bicarbonate, and sulfite and some of the ion pairs which exist in solution (magnesium sulfite, calcium bicarbonate, etc.). As stated earlier, the calcium salts of these compounds are relative insoluble. However, both magnesium and sodium salts are very soluble. High levels of dissolved magnesium (cations) require high levels of sulfite or other negative anions to remain in solution in order to maintain a neutral charge balance. A high level of sulfites increases the alkalinity of the scrubbing liquor, which permits a higher $SO_2$ removal rate per unit volume of slurry. Consequently, a lower liquor rate can be used. M.W. Kellogg is the primary vendor using magnesium promoted limestone systems, although other units use high magnesium limestone and benefit from this effect. This phenomena is not as effective in forced oxidized systems because there is a much lower concentration of sulfite ions in solution. Dissolved magnesium sulfate does not improve the $SO_2$ removal.

## Organic Acid Additives

The capability of carboxylic acids to improve $SO_2$ removal has been known for over 10 years, with initial research being performed by TVA (30). Other research performed by Rochelle (31), lead EPA to test adipic acid at the RTP laboratory facility (32), the Shawnee prototype unit (33), and at the commercial system of City Utilities (Southwest 1) (34). Adipic acid and the other carboxylic acids enhance $SO_2$ removal in a different manner than magnesium.

Adipic acid has dissociation constants at pH 5.4 and 4.4 which are near the operating range of most limestone systems. These are higher than the sulfurous acid dissociation constant so adipic can act as a base and neutralize the hydrogen ion generated during the absorption step.

$$SO_2(1) + H_2O \rightleftarrows H_2SO_3 \rightleftarrows H^+ + HSO_3^-$$

$$H^+ + HAd^- \rightleftarrows H_2Ad$$

$$H^+ + Ad^= \rightleftarrows HAd^-$$

This buffering capacity prevents the system pH from dropping, thereby maintaining a driving force for $SO_2$ removal.

Adipic acid also acts as a buffer for limestone dissolution, since it is a stronger acid than $H_2CO_3$. The protonated adipic species reacts with the dissolved carbonate ion to form water, $CO_2$, and the adipate ion.

$$CO_3^= + H_2Ad \rightleftarrows CO_2(g) + 2H^+ + Ad^=$$

The adipate ion is again available to assist in $SO_2$ removal. Since the adipic acid enhances limestone dissolution, a lower stoichiometric ratio can be used which reduces the limestone consumption and the resulting sludge volume. Also, since the adipate ion assists in $SO_2$ removal, forced oxidation of the sulfite ion does not impair the system removal efficiency as occurs in magnesium promoted systems which rely on high sulfite concentrations for liquid phase alkalinity. For the same reasons, high chloride levels also do not adversely affect the benefits of adipic acid.

The FGD system of City Utilities has added a feed system to permanently use dibasic acid (DBA) to control $SO_2$ emissions (35). Dibasic acid is a mixture of adipic, succinnic, and glutaric acids obtained during the production of adipic acid. It is a by-product stream and generally costs about half as much as pure adipic acid. The City Utilities installation is the first commercial-scale system to use an organic acid buffer. The Duck Creek unit of Central Illinois Light Company is also using DBA to improve their FGD system performance (36).

Absorber Design—As discussed previously, the major trend in absorber design has been away from packed contactors to open spray towers. This is due to the higher reliability and easier maintenance of the open vessels. Two new absorber designs have recently been evaluated which offer alternatives to the more conventional design. The designs are the cocurrent absorber, developed by EPRI and TVA, and the jet bubbling reactor, used in the Chiyoda 121 (CT-121) process. The CT-121 process is discussed in greater detail later in this section along with the jet bubbling reactor (JBR) design.

In the conventional spray tower design, the slurry and gas is contacted countercurrently with the gas entering near the bottom of the absorber. The cocurrent absorber has both the gas and liquid entering at the top of the absorber. The flue gas makes a 180° turn at the bottom of the absorber and then passes through a horzional mist eliminator. There are several advantages to this design. The mist eliminator and reheater are located on or near ground level which simplifies installation and maintenance. Also, the gas velocity can be much higher than in a countercurrent tower because of lower mist entrainment and a lower probability of flooding. This allows the absorber diameter to be smaller. In a comparison of the two absorber designs performed by Stearns-Roger for EPRI, the countercurrent absorber was

38′ in diameter while the equivalent capacity (125 MW module) cocurrent absorber had a 22′ diameter (11). A recently completed evaluation at TVA sponsored by EPRI and TVA showed the absorber to be capable of greater than 90 percent removal using limestone and several stages of grids. Some plugging problems were experienced with the mist eliminator at a gas velocity of 27 feet per second but did not occur at 20 fps (37, 38, 39, 40). Typical gas velocity in a countercurrent absorber is about 10 feet per second.

Several U.S. FGD systems vendor have expressed interest in the cocurrent design, however, none are yet offering it commercially (41). Cocurrent designs are in operation in Japan (42). The Merom units of Hoosier Energy built by Mitsubishi Heavy Industries are of cocurrent design, although the gas velocity is not as high as that tested at Shawnee.

### Economics—

The estimated and actual costs of limestone FGD systems have been reported for several years by many different groups. This section presents estimated costs for a common high sulfur application and variations of costs due to different design bases. The actual costs of commercial systems are presented as well as the results of an economic evaluation performed by City Utilities as part of their decision to use an organic acid to upgrade the Southwest 1 FGD unit.

Base Case Estimates—The two largest sponsors of FGD system cost estimate studies are the EPA-Industrial Environmental Research Laboratory, and the Electric Power Research Institute (EPRI). Most of the EPA's estimates have been performed by the Tennesse Valley Authority Office of Power (TVA). The Electric Power Research Institute has contracted with several A/E and consulting firms to perform estimates. Usually, all of the cost estimate reports include some type of base case limestone system. This system is designed to meet the applicable emission regulation using a high sulfur coal. Through the years, design and economic bases have changed in response to performance requirements and changing financial conditions.

Table 17 presents cost estimates for 11 new limestone FGD systems that were obtained from six reports. Although these estimates only range over a five-year period, changes in both capital and annual costs show over a hundred percent increase. Both design and economic factors are responsible for this increase. Figure 6 indicates graphically the impact of each factor for the differences in Estimates 1 and 5, the old and new base case system of TVA.

Estimates 1 and 2 by TVA compare the cost of meeting the 1971 NSPS or obtaining 90 percent removal of the base case 3.5 percent sulfur coal. As shown, the $SO_2$ removal increase from 79 to 90 percent alone does not significantly increase the cost. Estimate 3 is similar to Estimate 2, however, it incorporates an updated economic basis. These changes more than double the cost of the system. Estimates 4, 5 and 6 by TVA examine the spray tower system and an adipic acid enchanced system. Estimate 4, a natural oxidation system using a pond is more expensive than the TCA option, both in capital and operating costs (Estimate 3). Estimates 5 and 6 use forced oxidation and landfill disposal of the waste. The capital costs of the additional forced oxidation equipment is very small, however, the savings resulting from constructing a landfill rather than building a pond are significant. The operating costs of the landfill are higher than ponding, causing Estimate 5 to have a slightly higher revenue requirement than Estimate 4. Estimate 6 shows the benefit of using adipic acid to enhance the $SO_2$ removal. The absorber is designed for 95 percent removal, which allows bypassing of 6 percent of the flue gas, reducing the reheat costs. This estimate has significantly reduced capital and operating costs compared to the three previous systems.

Estimates 7 and 8, by Stearns-Roger for EPRI, used the 1979 NSPS and 1978 costs as bases. The costs are slightly lower than TVA's recent estimates. Although Estimate 8, with the cocurrent absorber has higher capital costs than the spray tower, higher contingency factors were used because no commercial cocurrent absorbers have as yet been demonstrated. The cocurrent design should result in equipment savings. In the cited study, total process capital costs for the spray tower and cocurrent systems were 117 and

TABLE 17. ESTIMATES OF NEW LIMESTONE FGD SYSTEMS

| Estimate Number | 1 | 2 | 3 |
|---|---|---|---|
| Source | TVA | TVA | TVA |
| Sponsor | EPA | EPA | EPA |
| Reference | 4 | 4 | 9 |
| Plant Size (MW) | 500 | 500 | 500 |
| Coal Type | Eastern | Eastern | Bituminous |
| Coal Sulfur (%) | 3.5 | 3.5 | 3.5 |
| NSPS | 1971 | 1979 | 1979 |
| Base Year - Capital | 1979 | 1979 | 1982 |
| Base Year - Operations | 1980 | 1980 | 1984 |
| Absorber Type | TCA | TCA | TCA |
| Absorber Removal (%) | 79 | 90 | 89 |
| Spare Module | No | No | Yes |
| L/G (gallons/1000 acf) | 50 | 50 | 58 |
| Additives | No | No | No |
| Sludge Disposal | Pond | Pond | Pond |
| Capital Cost ($/KWh) | 98 | 101 | 216 |
| Revenue Requirements (mills/KWh) | | | |
| First Year | 4.02 | 4.15 | 10.03 |
| Levelized | 5.16 | 5.33 | 13.81 |

(Continued)

TABLE 17. ESTIMATES OF NEW LIMESTONE FGD SYSTEMS (Continued)

| Estimate Number | 4 | 5 | 6 | 7 | 8 |
|---|---|---|---|---|---|
| Source | TVA | TVA | TVA | Stearns-Roger | Stearns-Roger |
| Sponsor | EPA | EPA | EPA | EPRI | EPRI |
| Reference | 9 | 9 | 9 | 12 | 12 |
| Plant Size (MW) | 500 | 500 | 500 | 500 | 500 |
| Coal Type | Bituminous | Bituminous | Bituminous | Bituminous | Bituminous |
| Coal Sulfur (%) | 3.5 | 3.5 | 3.5 | 4.0 | 4.0 |
| NSPS | 1979 | 1979 | 1979 | 1979 | 1979 |
| Base Year - Capital | 1982 | 1982 | 1982 | 1978 | 1978 |
| Base Year - Operations | 1984 | 1984 | 1984 | 1979 | 1979 |
| Absorber Type | Spray Tower | Spray Tower | Spray Tower | Spray Tower | Concurrent-Grid |
| Absorber Removal (%) | 89 | 89 | 95 | 90 | 90 |
| Spare Module | Yes | Yes | Yes | Yes | Yes |
| L/G (gallons/1000 acf) | 106 | 106 | 80 | 80 | 100 |
| Additives | No | No | Adipic Acid | No | No |
| Sludge Disposal | Pond | Landfill | Landfill | Landfill | Landfill |
| Capital Cost ($/KW) | 225 | 206 | 186 | 186 | 190 |
| Revenue Requirements (mills/KWh) | | | | | |
| First Year | 10.47 | 10.58 | 9.38 | 8.40 | 8.50 |
| Levelized | 14.42 | 15.08 | 13.31 | 12.94 | 13.18 |

(Continued)

TABLE 17. ESTIMATES OF NEW LIMESTONE FGD SYSTEMS (Continued)

| Estimate Number | 9 | 10 | 11 |
|---|---|---|---|
| Source | Bechtel | SRI | Beychok |
| Sponsor | EPRI | EPRI | EPRI |
| Reference | 10 | 11 | 17 |
| Plant Size (MW) | 2 x 500 | 500 | 500 |
| Coal Type | Bituminous | Eastern | Eastern |
| Coal Sulfur (%) | 4.0 | 4.0 | 3.5 |
| NSPS | 1978[a] | 1971[b] | 1971 |
| Base Year - Capital | 1978 | 1979 | 1977 |
| Base Year - Operations | 1978 | 1979 | 1977 |
| Absorber Type | Spray Tower | Grid | Spray Tower |
| Absorber Removal (%) | 85 | 84 | 90 |
| Spare Module | Yes | Yes | Yes |
| L/G (gallons/1000 acf) | 100 | 145 | -- |
| Additives | No | No | No |
| Sludge Disposal | Landfill | Landfill | Landfill |
| Capital Cost ($/KW) | 157 | 165 | 95 |
| Revenue Requirements (mills/KWh) | | | |
| First Year | 8.89 | 7.35 | 6.61 |
| Levelized | 12.7 | 12.97 | 10.00 |

[a] 1978 Proposed NSPS.

[b] Design basis was similar to 1979 NSPS requirements.

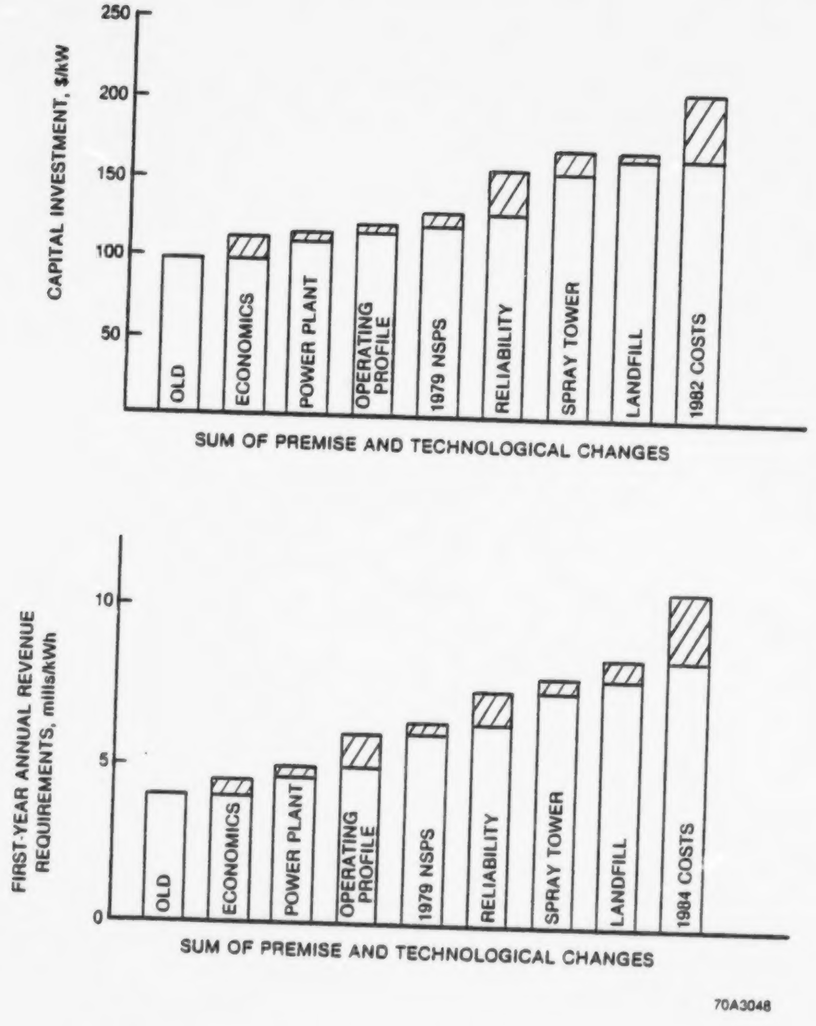


Figure 6. Stepwise Conversion of TVA's Limestone Process Costs From Old to New Premises and Technology

112 dollars per KW, respectively. Estimates 9 and 10, by Bechtel and SRI, have design requirements similar to the current NSPS and are expressed in 1978 and 1979 dollars. Their costs are close to those shown in Figure 4 before the 1982 costs are included. Estimate 11 by Beychok is based on the 1971 NSPS and is very close to Estimate 1 by TVA.

The wide range of estimates, which are generally only accurate to ±30 percent, makes it difficult to compare design differences or economic basis used in the various reports. However, when these estimates are examined closely and adjusted to a common basis they agree reasonably well. The more recent estimates indicate that a new limestone FGD system for a high sulfur coal should cost about $200 per kilowatt. The next section examines variations in costs due to changes in specific design parameters.

Parametric Case Estimates—A recent report has been prepared by TVA which presents costs for a variety of specific FGD applications (8). These costs were estimated using the Shawnee lime—limestone computer program (43). Therefore, they are all on a consistent basis and reflect differences in the system cost due to design factors rather than variations in different estimation techniques. Although the estimates are preliminary grade economics, the difference in the estimates are thought to be real. All of the costs are for new systems regulated by the 1979 NSPS.

Most of the following estimates are made relative to the base case study; Estimate 5 in Table 17 which is a spray tower FGD system on a 500 MW unit burning 3.5 percent sulfur coal. Table 18 presents the costs of this same type of system for four different coals at three power plant sizes. The low sulfur case is for a subbituminous coal, the other three coals are bituminous. As can be seen, there is an increasing capital and annual revenue requirement with higher sulfur levels. The unit cost also increases as the plant size drops. Both of these trends are expected, reflecting the economy of scale of larger units, and the increased reagent consumption and associated slurry processing at higher sulfur levels.

TABLE 18. ECONOMICS OF THE LIMESTONE FGD SYSTEM AS A FUNCTION OF POWER UNIT SIZE AND COAL SULFUR LEVEL (Spray tower, forced oxidation, landfill, 1979 NSPS)

| Power unit size (MW) | Sulfur Content (%) | | | |
|---|---|---|---|---|
| | 0.7 | 2.0 | 3.5 | 5.0 |
| **Capital Investment ($/KW)** | | | | |
| 200 | 220 | 260 | 295 | 313 |
| 500 | 141 | 173 | 206 | 218 |
| 1,000 | 109 | 152 | 177 | 187 |
| **First Year Revenue Requirements (mills/KWh)** | | | | |
| 200 | 11.15 | 13.40 | 15.21 | 16.17 |
| 500 | 6.84 | 8.80 | 10.58 | 11.36 |
| 1,000 | 5.15 | 7.47 | 8.97 | 9.63 |
| **Levelized Revenue Requirements (mills/KWh)** | | | | |
| 200 | 15.76 | 19.10 | 21.69 | 23.08 |
| 500 | 9.56 | 12.51 | 15.08 | 16.25 |
| 1,000 | 7.13 | 10.50 | 12.71 | 13.73 |

Source: Reference 9.

Table 19 shows the costs of the 500 MW, 3.5 percent sulfur coal system as a function of the absorber type and waste disposal. Here, the TCA is shown to have lower capital and annual costs than the spray tower and the venturi-spray tower. This estimation may be erroneous because of constraints within the computer model. The TCA design has a maximum gas velocity of 12.5 fps whereas the spray tower is limited to 10 fps. In identical applications, the spray tower diameter is larger, making the system more costly. Since most of the indirect capital charges are assumed to be constant percentages of the direct capital costs, the total cost is also higher for the spray tower. This same phenomena occurs with the operating costs. In this case, the maintenance costs are assumed to be proportional to the capital costs. However, operating experience with TCA's has shown them to require a greater level of maintenance than spray towers. Thus, the annual maintenance cost for the TCA system is probably underestimated relative to the spray tower unit.

The costs of using forced oxidation and a landfill or natural oxidation and a pond are also shown in Table 19. In all cases, the capital costs were several percent higher for the pond cases due to the extra construction involved. Operating costs were 3-5 percent lower for the pond cases because unlike the landfill, personnel were not required on a daily basis to tend the pond. One important factor which could change these results is the distance from the FGD systems to the disposal site. In both cases, the pond or landfill is assumed to be one mile from the process. If greater distances are involved, due to unavailability or expensive land costs, the landfill operation could have lower operating costs than ponding. In the pond applications, supernatant water is pumped back to the FGD system.

Another variation examined by TVA was the cost of using additives (such as magnesium oxide, adipic acid, and DBA) to enhance the $SO_2$ removal. These estimates are shown in Table 20. The lower capital cost for the buffered systems are due to the lower L/G required, smaller limestone preparation equipment (due to higher utilization) and the use of bypass for some reheat. Operating costs are lower because of lower capital charges and lower energy and raw material requirements. The worst case adipic acid cost estimate

TABLE 19. ECONOMICS OF THE LIMESTONE SYSTEM AS A FUNCTION OF ABSORBER TYPE AND WASTE DISPOSAL (500 MW, 1979 NSPS, 3.5 Percent Sulfur)

| | Capital Investment ($/KW) | Revenue Requirement (mills/KWh) First Year | Revenue Requirement (mills/KWh) Levelized |
|---|---|---|---|
| Forced Oxidation, Landfill | | | |
| TCA | 197 | 10.09 | 14.37 |
| Spray Tower | 206 | 10.58 | 15.08 |
| Venturi-Spray Tower | 218 | 11.20 | 15.96 |
| Natural Oxidation, Pond | | | |
| TCA | 216 | 10.03 | 13.81 |
| Spray Tower | 225 | 10.47 | 14.42 |
| Venturi-Spray Tower | 243 | 11.37 | 15.68 |

Source: Reference 9.

TABLE 20. ECONOMICS OF THE LIMESTONE SYSTEM WITH ADDITIVES (500 MW, Spray tower, forced oxidation, 3.5 percent sulfur, 1979 NSPS)

| Case | Capital Investment ($/KW) | Revenue Requirement (mills/KWh) | |
|---|---|---|---|
| | | First Year | Levelized |
| Base | 206 | 10.58 | 15.08 |
| Magnesium Oxide | 205 | 10.49 | 14.93 |
| Adipic Acid | 186 | 9.38 | 13.31 |
| DBA | 186 | 9.35 | 13.25 |
| Worst Case Adipic Acid | 186 | 9.59 | 13.69 |

Source: References 9 and 44.

assumes a degradation factor of 9 (instead of 3) and an adipic acid cost of $1,800 per ton (instead of $1,500). The degradation factor is the ratio of the actual addition rate to the theoretical rate. Adipic acid is lost in liquid leaving the system and also chemically degraded. As can be seen, these worst case assumptions do not change the cost savings of the buffered systems relative to the base case system. The use of adipic acid would reduce the cost of the FGD system by 146 million dollars over the 30 year life of the plant (165,000 hours, 500 MW). The use of magnesium oxide does not significantly reduce the capital or operating costs because the limestone stoichiometry is not reduced. Although the L/G ratio is lowered slightly, the cost of magnesium oxide offsets this benefit.

Costs of Existing Limestone Systems—The PEDCo survey of commercial FGD systems contains operating and capital costs of many FGD systems. The cost reported by the utility is presented along with an adjusted cost in Table 21. The adjusted costs are presented in mid-1981 dollars. Several factors are adjusted in order to put the costs on a similar basis. For instance, the particulate collection costs are deducted whereas any specific modifications required for the FGD system are included. A 65 percent capacity factor is used to obtain annual unit costs.

Table 21 presents the reported and adjusted costs of the limestone FGD systems in the PEDCo survey. The average adjusted capital cost is $98 per KW which is fairly low compared to the estimates previously shown. However, several factors need to be considered in order to make a fair comparison. In the PEDCo adjustment procedure, the FGD system cost is divided by the gross generating capacity rather than the equivalent scrubbed gas capacity. Quite a few of the systems with low dollar per KW values only scrub 50-70 percent of the flue gas (Tombigbee 2, 3; Apache 2, 3; Cholla 1; Craig 1, 2; Coronado 1, 2; Morrow 1, 2). The third column reflects the cost of the FGD systems based on the amount of flue gas scrubbed. The equivalent scrubbed capacity costs show an average capital cost of $120 per KW in 1981 dollars.

TABLE 21. OPERATIONAL LIMESTONE SYSTEM COSTS

| Unit | Capital costs ($/KW) | | | Annual costs (mills/KWh) | |
|---|---|---|---|---|---|
| | Reported | Adjusted | ESC[a] | Reported | Adjusted |
| Tombigbee 2 | 27.4 | 38.3 | 54.6 | 0.33 | 3.18 |
| Tombigbee 3 | 27.4 | 38.3 | 54.6 | 0.33 | 3.18 |
| Apache 2 | 41.8 | 67.5 | 134.3 | 0.10 | 3.75 |
| Apache 3 | 37.0 | 59.8 | 119.0 | 0.10 | 3.75 |
| Cholla 1 | 52.0 | 81.3 | 81.3 | – | 4.79 |
| Cholla 2 | 168.0 | 162.1 | 162.1 | 0.75 | 8.22 |
| Duck Creek 1[b] | 74.8 | 132.2 | 132.2 | 5.04 | 5.78 |
| Craig 1 | 86.5 | 105.9 | 117.5 | 0.53 | 5.33 |
| Craig 2 | 73.5 | 90.3 | 100.2 | 0.48 | 4.56 |
| Petersburg 3[b] | 104.9 | 162.1 | 162.1 | – | 9.74 |
| Lacygne 1[b] | 53.7 | 100.1 | 100.1 | 5.38 | 11.29 |
| Jeffery 1 | 37.9 | 55.9 | 74.5 | 0.11 | 2.42 |
| Jeffery 2 | 37.8 | 41.2 | 58.9 | 0.29 | 1.65 |
| Coronado 1 | 79.4 | 86.5 | 108.1 | 0.85 | 7.70 |
| Coronado 2 | 79.4 | 86.5 | 108.1 | 0.85 | 7.70 |
| Winyah 2 | 23.7 | 47.0 | 94.0 | 0.29 | 1.84 |
| Winyah 3 | 38.4 | 66.6 | 66.6 | – | 4.83 |
| Morrow 1 | 53.7 | 118.5 | 191.1 | – | 6.94 |
| Morrow 2 | 53.7 | 118.5 | 191.1 | – | 6.94 |
| Marion 4[b] | 87.9 | 120.8 | 120.8 | 0.98 | 7.75 |
| Southwest 1[b] | 86.3 | 143.4 | 143.4 | 1.20 | 8.21 |
| Dallman 3[b] | 170.4 | 194.3 | 215.3 | 4.68 | 14.56 |
| Widdows Creek 8[b] | 87.1 | 158.1 | 158.1 | 7.76 | 7.28 |
| Average | 68.8 | 98.9 | 119.5 | 1.67 | 6.15 |
| Std. Dev. | 39.4 | 44.9 | 45.5 | 2.32 | 3.14 |

[a] Equivalent scrubbed capacity.

[b] Sulfur content above 1.5 percent.

Source: Reference 2.

The design removal of these units varies considerably. None of these systems were built under the present NSPS. Additionally, only 7 of the 23 systems burn coal with a sulfur content above 1.5 percent. The average adjusted capital cost of these seven units is $149 per KW in mid-1981 dollars. None of these seven units have spare modules, which would increase the cost by 20-30 percent. Increasing the base year from 1981 to 1982 would raise the average cost to just about $200 per KW which is approximately the same cost predicted in the TVA study. A recent paper by Baasel examines the differences in reported and actual limestone FGD system costs. The results indicate that the estimates and actual costs agree very well when placed on a common basis (45).

Commercial DBA Retrofit--A commercial-scale demonstration program of organic acid additives to a limestone FGD system has recently been completed at City Utilities (Springield, Missouri) Southwest 1 power plant (34, 35, 46). Both adipic and dibasic acid were tested in a 9-month program sponsored by the EPA.

The City Utilities system uses two turbulent contact absorbers (TCA) for the 194 MW unit. During the course of the test program, City Utilities changed one of the absorbers to a tray tower in hopes of improving the system reliability. At the end of the test program, the second absorber was also converted to a tray tower. In order to achieve continuous compliance with the 1971 NSPS, City Utilities examined four alternatives to increase the systems's $SO_2$ removal efficiency. These options were: (1) adipic acid addition to the tray towers, (2) DBA addition to the tray towers, (3) increased L/G in the tray towers, and (4) conversion to spray towers at a higher L/G. Option 4 was eliminated due to technical concerns about the high gas velocity in the absorbers (35).

A present worth analysis was conducted of the remaining three options which showed the DBA addition would be economical choice. Over a 15-year time span, the DBA option had a present value cost of 1.6 million dollars.

The adipic acid option cost 2.35 million. The increased L/G option was estimted at having a 2 million dollar present worth. This corresponds to approximately a 0.05 mill per KWh saving by using DBA instead of adipic. It is interesting to note that TVA has predicted a 0.06 mill per KWh savings (See Table 20) for these two cases when applied to a new system. Although the unit costs are very small, the cost differential over the life of the plant is fairly substantial. City Utilities is now using DBA addition to remain in compliance with their $SO_2$ emission regulation. The use of DBA has also allowed the utility to burn cheaper, higher sulfur coal and still remain in compliance while achieving a significant fuel cost savings (35).

## Lime

Wet lime systems are the second largest category of FGD systems, however future applications are on the decline. Lime systems account for 30 percent of the operating FGD capacity. However, only 14 percent of the units planned, contracted, or under construction will use the wet lime process. Table 22 lists the operational and regulatory status of the 32 commercial lime FGD systems. As can be seen in Table 22, over two-thirds of the lime systems are operational. In contrast, only 27 percent of the limestone systems previously listed in Table 14 have been commissioned.

The higher reagent cost of lime and subsequently higher annualized cost is the main reason that lime systems have not grown significantly in number. Although most early high sulfur systems selected lime because of its higher reactivity, better designs of limestone systems have negated this advantage.

### Process Description—

Chemically, lime systems are only slightly different from limestone systems. Limestone systems desorb $CO_2$ in the scrubber, where lime systems absorb a small amount of $CO_2$. Consequently, lime is generally more reactive than limestone as a reagent. The process design is also only moderately

TABLE 22. OPERATIONAL AND REGULATORY STATUS OF LIME FGD SYSTEMS AS OF DECEMBER 1981

| Applicable $SO_2$ Standard | MW Capacity (Number of units) | | | | |
|---|---|---|---|---|---|
| | Operating | In Construction | Contracted | Planned | Total |
| 1979 NSPS | - | 65 (1) | 650 (1) | 330 (1) | 1045 (3) |
| 1971 NSPS | 3,556 (8) | 995 (2) | - | - | 4,551 (10) |
| >1979 NSPS | - | 1,810 (3) | - | - | 1,810 (3) |
| >1971 NSPS, <1979 NSPS | 5,486 (12) | - | - | - | 5,486 (12) |
| ≤1971 NSPS | 844 (4) | - | - | - | 844 (4) |
| TOTAL | 9,886 (24) | 2,870 (6) | 650 (1) | 330 (1) | 13,736 (32) |

Source: Reference 2.

different, the major distinction being the reagent preparation step. Figure 7 is a flow diagram of a typical lime process. Pebble lime is usually delivered to the plant and stored in silos. Several plants receive lime slurry directly; notably the units of Louisville Gas and Electric which use carbide lime. Their systems will be discussed in more detail later.

The delivered lime (often called quicklime—CaO) must be hydrated to calcium hydroxide. A slaker is used to perform this reaction, which is highly exothermic. Temperature control is important in maintaining good lime reactivity. The slurry is fed to the reaction tank to maintain the desired pH, usually between 7 and 9. Above pH 9, calcium carbonate scaling in the upper parts of the absorber can be a problem. Below pH 7, the $SO_2$ removal generally is unsatisfactory. The reactions which occur are similar to those in a limestone system.

$$SO_2(g) + Ca(OH)_2(l) \rightleftarrows CaSO_3 \cdot 1/2H_2O(s) + 1/2H_2O$$

$$SO_3^{=} + 1/2O_2(g) \rightarrow SO_4^{=}$$

$$Ca^{++} + SO_4^{=} + 2H_2O \rightleftarrows CaSO_4 \cdot 2H_2O(s)$$

Although lime is not very soluble at the lime slurry pH of 12, it dissolves readily at the more neutral conditions in the FGD system. This results in good utilization of the lime and quick pH response to changes in the lime addition rate, unlike limestone systems which are heavily buffered by the carbonate ion. Consequently, the liquid to gas ratio can be lower, which reduces the pumping horsepower while achieving the desired $SO_2$ removal.

The same $SO_2$ removal, oxidation, and precipitation reactions occur as in limestone systems, except that a water molecule is liberated from the calcium hydroxide rather than $CO_2$ when $SO_2$ neutralization occurs. The solids are thickened and filtered before disposal and present similar disposal problems as discussed previously for limestone systems.

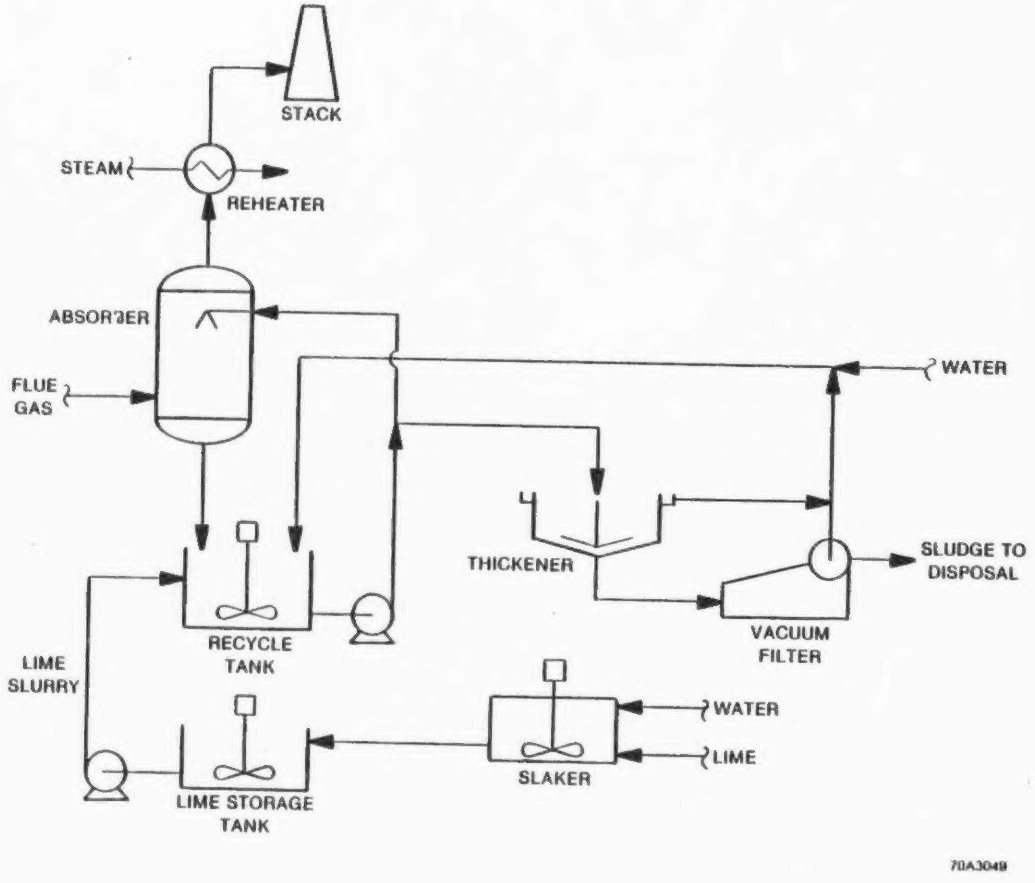

STACK
STEAM
REHEATER
ABSORBER
FLUE GAS
WATER
THICKENER
SLUDGE TO DISPOSAL
VACUUM FILTER
RECYCLE TANK
LIME SLURRY
WATER
LIME
SLAKER
LIME STORAGE TANK
70A3049

Figure 7. Lime Process Flow Diagram

Commercial Designs--

Table 23 presents selected design features of the lime systems identified in the FGD data base. Several observations can be made comparing the design of lime and limestone systems.

The L/G ratio of lime and limestone units regulated by the 1971 NSPS are similar. It averages 60 for lime systems and 63 for limestone systems. Also, the percent sulfur in the coal for boilers using FGD subject to the 1971 NSPS is not statistically different; 2.57 percent for lime and 2.14 percent for limestone. Two factors which are different are the absorber design and the number of retrofit units.

Table 24 lists the sequential order of absorber types for lime systems built by the various vendors. There is a greater balance in the type of absorbers used as opposed to limestone systems. Intially, packed towers or venturis were offered by the vendors, whereas later units are primarily spray towers.

Table 25 compares the early applications of absorber types for both lime and limestone systems. As can be seen, lime systems were equal in number to limestone units through 1977. Since that time, however, limestone units have been in the majority. It is most probable that unfavorable maintainence experiences with packed towers on lime systems hastened the acceptance of spray towers for both processes.

Another interesting observation from Table 25 is the number of retrofit systems which used lime instead of limestone. Eight of the first eleven lime systems were retrofitted whereas only three of the first nine limestone systems (not including Shawnee) were retrofit through 1977. Three of these eight retrofit lime units are owned by Louisville Gas and Electric which uses carbide lime as the alkali. This lime is available nearly as a waste by-product from the manufacturer of acetylene and is obtained at a competitive price. The other five systems slake lime on site. Perhaps early

TABLE 23. LIME FGD SYSTEMS DESIGNS

| Plant | Startup | Percent Sulfur | L/G (gal/1000 acf) | Absorber Removal (%) | Unit Removal (%) | Absorber Type[a] | Vendor[b] |
|---|---|---|---|---|---|---|---|
| | | | 1979 NSPS | | | | |
| J.K. Smith 1 | 08/87 | 1.5 | – | – | – | – | BW |
| J.B. Sims 3 | 01/83 | 2.8 | 110 | 90 | 90 | ST | BW |
| | | | 1971 NSPS | | | | |
| Green 1 | 12/79 | 3.8 | 58 | 90 | 90 | ST | AAF |
| Green 2 | 11/80 | 3.8 | 58 | 90 | 90 | ST | AAF |
| East Bend 2 | 03/81 | 3.0 | 62 | 87 | 87 | T | BW |
| Spurlock 2 | 01/83 | 3.5 | – | 90 | 90 | ST | THY |
| Mill Creek 4 | 06/82 | 3.8 | – | 85 | 85 | P | AAF |
| Pleasants 1 | 03/79 | 3.0 | 70[c] | 90 | 90 | T | BW |
| Pleasants 2 | 10/80 | 3.0 | 70[c] | 90 | 90 | T | BW |
| Hunter 1 | 05/79 | 0.6 | 52 | 80 | 80 | ST | GE |
| Hunter 2 | 06/80 | 0.6 | 52 | 80 | 80 | ST | GE |
| Huntington 1 | 05/78 | 0.6 | 52 | 80 | 80 | ST | GE |
| | | | Stricter Standard Than 1979 NSPS | | | | |
| Four Corners 4 | 12/84 | 0.8 | – | 68 | – | ST | BW |
| Four Corners 5 | 12/84 | 0.8 | – | 68 | – | ST | BW |
| Mitchell 33 | 08/82 | 2.8 | 110 | 95 | 95 | ST | GE |
| | | | Stricter Standard Than 1971 NSPS But Less Than 1979 NSPS | | | | |
| Conesville 5 | 01/77 | 4.7 | 62 | 90 | 90 | P | UOP |
| Conesville 6 | 06/78 | 4.7 | 62 | 90 | 90 | P | UOP |
| Elrama 1-4 | 10/75 | 2.2 | 39 | 83 | 83 | V | GE |
| Phillips 1-6 | 07/73 | 1.9 | 42 | 83 | 83 | V | GE |
| Hawthorn 3 | 11/72 | 0.6 | 26 | 70 | 70 | P | CE |
| Hawthorn 4 | 08/72 | 0.6 | 26 | 70 | 70 | P | CE |
| Cane Run 4 | 08/76 | 3.8 | 48 | 85 | 85 | P | AAF |
| Cane Run 5 | 12/77 | 3.8 | 61 | 85 | 85 | ST | CE |
| Mill Creek 3 | 08/78 | 3.8 | 71[c] | 85 | 85 | P | AAF |
| Bruce Mansfield 1 | 12/75 | 3.0 | 50 | 92 | 92 | V | GE |
| Bruce Mansfield 2 | 07/77 | 3.0 | 50 | 92 | 92 | V | GE |
| Bruce Mansfield 3 | 06/80 | 3.0 | – | 92 | 92 | ST | K |

(Continued)

TABLE 23. LIME FGD SYSTEMS DESIGN (Continued)

| Plant | Startup | Percent Sulfur | L/G (gal/1000 acf) | Absorber Removal (%) | Unit Removal (%) | Absorber Type[a] | Vendor[b] |
|---|---|---|---|---|---|---|---|
| | | | Standard Equal to or Less Than 1971 NSPS | | | | |
| Green River 1-3 | 09/75 | 4.0 | 39[c] | 80 | 80 | V-P | AAF |
| Mill Creek 1 | 12/80 | 3.8 | 123 | 87 | 85 | ST | CE |
| Mill Creek 2 | 12/81 | 3.8 | 123 | 87 | 85 | ST | CE |
| Paddy's Run 6 | 04/73 | 2.5 | 64 | 90 | 90 | P | CE |

[a] Absorber Type: P - packed tower; ST - spray tower; T - tray tower; V - venturi.

[b] Vendors: AAF - American Air Filter; BW - Babcock and Wilcox; CE - Combustion Engineering;
GE - General Electric; K - Kellogg; THY - Thyssen/CEA;
UOP - Air Correction Division.

[c] Includes venturi or quencher L/G.

Source: Reference 2.

TABLE 24. SEQUENTIAL ORDER OF LIME ABSORBERS

| Vendor | Absorber Types[a] | Total |
|---|---|---|
| American Air Filter | V-P, P(2), ST(2), P | 6 |
| Babcock and Wilcox | T(3), ST(3), * | 7 |
| Combustion Engineering | P(3), ST(3) | 6 |
| General Electric | V(4), ST(4) | 8 |
| Kellogg | ST | 1 |
| Thyssen/CEA | * | 1 |
| UOP | P(2) | 2 |
| Undetermined | ST | 1 |
|  |  | 32 |

[a] Absorber Type: P - packed; ST - spray tower; T - tray; V - venturi; * - unknown.

Source: Reference 2.

## TABLE 25. ABSORBER TYPES BY SYSTEM AND YEAR[a]

| Year | Lime Systems |  | Limestone Systems |  |
|---|---|---|---|---|
|  | New | Retrofit | New | Retrofit |
| 1972 |  | P(2), P[b] |  | V-ST[b] |
| 1973 |  | V,P | V-T | G |
| 1974 |  |  |  |  |
| 1975 | V | V,V-P |  |  |
| 1976 |  | P | G |  |
| 1977 | P,V | ST | G,P(2),V-T | V-P,V-ST |
| 1978 | ST | P(2) | G(4),ST(3),<br>V-ST |  |
| 1979 | ST(2),T |  | G(3),ST(3), |  |

[a] Absorber Types: G - grid; P - packed; ST - spray tower; T - tray; v - venturi.

[b] Shawnee prototype units.

Source: Reference 1.

operating difficulties with limestone systems caused many of the utilities to favor lime usage. But, beginning in 1977, limestone systems started being the systems of choice for new units.

Recent Innovations—

Only six lime systems are scheduled to start up in the 1982-84 time period. The majority of these systems will be spray towers. No other recent process improvements obtained with limestone systems will be employed.

TVA has performed limited research with adipic acid and forced oxidation at the Shawnee prototype systems using lime. Tests have been conducted on the venturi, the venturi-spray tower, and the TCA systems in both natural and forced oxidation modes. Adipic acid has shown the same capability for increasing $SO_2$ removal as demonstrated with limestone systems (47, 48). The test series demonstrating forced oxidation within a single loop TCA system is most applicable for improving $SO_2$ removal and solids handling properties of the typical lime system. Magnesium oxide addition to lime systems has also been shown to increase the $SO_2$ removal efficiency (49).

Forced oxidation is typically not performed in lime systems because the reaction tank pH is above the optimum oxidation level. If the pH is lowered in order to increase the oxidation, the $SO_2$ removal decreases. Unlike limestone systems which contains a large solid phase reservoir of carbonate alkalinity, in lime systems once the sulfite is oxidized to sulfate, the available alkalinity is decreased. Adipic acid addition in a lime system provides sufficient alkalinity at the lower pH to maintain good $SO_2$ removal. The sulfite ions can then be oxidized without reducing the removal efficiency.

The other advantage of adipic acid use in limestone systems, improved reagent utilization, is not as large a factor in lime systems. This is because lime is inherently more reactive and usually requires as a maximum a

10 percent excess. Limestone systems typically operate at a 30-50 percent excess addition rate which can be lowered to 10-20 percent excess by using organic acids.

Economics—

In general, the capital costs of lime systems are slightly lower than limestone systems because on a weight basis, fewer tons of lime are required for a given $SO_2$ removal rate. Also, lime is more reactive which reduces the pumping and absorber size requirements. However, the higher cost of lime often results in higher annualized costs. This section presents base case costs, parametric studies, and the actual and adjusted costs of commercial lime systems.

Base Case Estimates—Estimated costs of new lime systems for the base case 500 MW high sulfur system are presented in Table 26. These studies were performed by TVA and Bechtel, for the EPA and EPRI respectively.

Estimates 1 and 2 by TVA are directly comparable to the limestone system costs of Table 17, Estimates 1 and 2. As can be seen, the capital costs are about 8 percent lower for the lime systems although the annual costs are 6-8 percent higher. Estimate 3, by Bechtel, is comparable to Estimate 9 in Table 17. Again, the same capital and annual cost trends are evident. Estimate 3 is higher than 1 and 2 due to the extra module, bypass ducts around the system, and larger reagent preparation facilities. Estimate 4 is for a western coal application. The similar limestone system (not shown in Table 17) has capital, annual, and levelized costs of $118 per KW; 5.77 and 8.0 mills per KWh (10). The lime system is estimated to be more economical than limestone when only a small amount of lime is required. This trend will be discussed further in the next section.

Estimates 5 and 6 are from a recent TVA study using their updated design and economic bases. Estimate 5 is similar in design to Estimate 1.

## TABLE 26. COST ESTIMATES FOR NEW LIME FGD SYSTEMS

| Estimate Number | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| Source | TVA | TVA | Bechtel | Bechtel | TVA | TVA |
| Sponsor | EPA | EPA | EPRI | EPRI | EPA | EPA |
| Reference | 4 | 4 | 10 | 10 | 9 | 9 |
| Plant size (MW) | 500 | 500 | 2 x 500 | 2 x 500 | 500 | 500 |
| Coal type | Eastern | Eastern | Bituminous | Subbituminous | Eastern | Eastern |
| Coal sulfur (%) | 3.5 | 3.5 | 4.0 | 0.48 | 3.5 | 3.5 |
| NSPS | 1971 | 1979 | 1978[a] | 1978[a] | 1979 | 1979 |
| Base year - capital | 1979 | 1979 | 1978 | 1978 | 1982 | 1982 |
| Base year - operations | 1980 | 1980 | 1978 | 1978 | 1984 | 1984 |
| Absorber type | TCA | TCA | Spray Tower | Spray Tower | TCA | Spray Tower |
| Absorber removal (%) | 79 | 90 | 85 | 85 | 89 | 89 |
| Spare module | No | No | Yes | Yes | Yes | Yes |
| L/G (gallons/1000 acf) | 45 | 45 | 85 | 55 | 50 | 70 |
| Sludge disposal | Pond | Pond | Landfill | Landfill | Pond | Landfill |
| Capital Cost ($/KW) | 90 | 94 | 149 | 110 | 192 | 178 |
| Revenue Requirements (Mills/KWh) | | | | | | |
| First year | 4.25 | 4.46 | 9.0 | 5.46 | 10.19 | 10.35 |
| Levelized | 5.31 | 5.54 | 13.2 | 7.6 | 14.66 | 15.31 |

[a] Proposed 1978 NSPS.

Estimate 6 uses forced oxidation and landfill disposal to obtain lower capital costs; however, the landfill labor costs increase the annualized charges to above the pond disposal case.

Parametric Case Estimates—A recent TVA study (8) presents the economics of several different applications of lime systems. Table 27 presents the results of this study. Cases 1-6 are for natural oxidation, pond disposal applications at various sulfur levels and absorber designs. The lowest costs are shown at the lowest sulfur level. The TCA system also has a slightly lower cost than the spray tower.

Cases 7-9 use forced oxidation and landfill disposal of waste sludge. In each instance, the capital costs are lower while the levelized costs are higher for the landfill cases compared to the equivalent pond disposal case. Cases 10-12 show that magnesium addition does not provide an economic benefit to lime systems. All of the costs are slightly higher than the corresponding Cases 7-9.

Table 28 compares selected lime and limestone system costs. In all cases, the lime systems have lower capital costs, reflecting the reduced reagent volume and the higher reactivity of lime. However, for all but the two low sulfur cases, the levelized revenue requirements for the limestone systems are lower than the lime units. This is due to the lime reagent costs, which are 3.7 times greater than the limestone costs. At the low sulfur level however, this is not a major portion of the levelized cost. Other studies have shown that at small plant sizes (200 MW or less), lime systems are also competitive due to the reduced reagent volume (4, 50).

Costs of Existing Lime Systems—Reported and adjusted capital and operating costs of commercial lime systems are presented in Table 29. The average adjusted cost is $117 per KW with an 8.1 mill per KWh annual cost. For the five low sulfur systems the average capital and annual costs are $72 per KW and 5.1 mills per KWh. It is interesting to note that the 16 low sulfur limestone systems in Table 21 averaged $79 per KW and 4.8 mills per

## TABLE 27. PARAMETRIC LIME SYSTEM ECONOMICS

(500 MW, 1979 NSPS)

| Case | Oxidation | Sludge Disposal | Coal Sulfur % | Absorber Type | Capital Investment ($/KW) | Revenue Requirements (mills/KWh) First Year | Revenue Requirements (mills/KWh) Levelized |
|---|---|---|---|---|---|---|---|
| 1 | Natural | Pond | 0.7 | ST | 121 | 5.56 | 7.63 |
| 2 | Natural | Pond | 2.0 | ST | 159 | 7.92 | 11.18 |
| 3 | Natural | Pond | 3.5 | ST | 192 | 10.19 | 14.66 |
| 4 | Natural | Pond | 5.0 | ST | 211 | 11.51 | 16.71 |
| 5 | Natural | Pond | 3.5 | TCA | 189 | 10.09 | 14.55 |
| 6 | Natural | Pond | 3.5 | V-ST | 206 | 10.88 | 15.63 |
| 7 | Forced | Landfill | 3.5 | ST | 178 | 10.35 | 15.31 |
| 8 | Forced | Landfill | 3.5 | TCA | 172 | 10.16 | 15.07 |
| 9 | Forced | Landfill | 3.5 | V-ST | 185 | 10.77 | 15.92 |
| 10[a] | Forced | Landfill | 3.5 | ST | 179 | 10.40 | 15.37 |
| 11[a] | Forced | Landfill | 3.5 | TCA | 181 | 10.45 | 15.43 |
| 12[a] | Forced | Landfill | 3.5 | V-ST | 195 | 11.00 | 16.18 |

[a] Magnesium oxide addition.

Source: Reference 9.

TABLE 28. COMPARISON OF LIME AND LIMESTONE SYSTEM COSTS

(500 MW, 1979 NSPS)

| System | Coal Sulfur % | Waste Disposal | Absorber Type | Capital Cost ($/KW) | Revenue Requirements (mills/KWh) | |
|---|---|---|---|---|---|---|
| | | | | | First Year | Levelized |
| Limestone | 0.7 | Pond | ST | 147 | 6.56 | 8.91 |
| Lime | 0.7 | Pond | ST | 121 | 5.56 | 7.62 |
| Limestone | 2.0 | Pond | ST | 183 | 8.46 | 11.61 |
| Lime | 2.0 | Pond | ST | 159 | 7.92 | 11.18 |
| Limestone | 3.5 | Pond | ST | 225 | 10.47 | 14.42 |
| Lime | 3.5 | Pond | ST | 192 | 10.19 | 14.66 |
| Limestone | 5.0 | Pond | ST | 247 | 11.44 | 15.73 |
| Lime | 5.0 | Pond | ST | 211 | 11.51 | 16.71 |
| Limestone | 3.5 | Landfill | ST | 206 | 10.58 | 15.08 |
| Lime | 3.5 | Landfill | ST | 178 | 10.35 | 15.31 |
| Limestone | 3.5 | Landfill | TCA | 197 | 10.09 | 14.37 |
| Lime | 3.5 | Landfill | TCA | 172 | 10.16 | 15.07 |
| Limestone | 3.5 | Pond | TCA | 216 | 10.03 | 13.81 |
| Lime | 3.5 | Pond | TCA | 189 | 10.09 | 14.55 |

Source: Reference 9.

TABLE 29. OPERATIONAL LIME SYSTEM COSTS

| Unit | Capital costs ($/KW) Reported | Capital costs ($/KW) Adjusted | Annual costs (mills/KWh) Reported | Annual costs (mills/KWh) Adjusted |
|---|---|---|---|---|
| Conesville 5 | 50.3 | 99.4 | 5.52 | 6.78 |
| Conesville 6 | 50.3 | 99.4 | 5.52 | 6.78 |
| Elrama 1-4 | 120.2 | 187.8 | 7.18 | 12.89 |
| Phillips 1-6 | 127.0 | 210.0 | 11.0 | 17.65 |
| Hawthorn 3* | 29.4 | 62.8 | 1.15 | 5.25 |
| Hawthorn 4* | 29.4 | 62.8 | 1.15 | 5.25 |
| Green River 1-3 | 62.0 | 117.8 | 5.20 | 10.99 |
| Cane Run 4 | 66.6 | 115.2 | 1.29 | 6.20 |
| Cane Run 5 | 68.8 | 102.4 | 0.92 | 5.30 |
| Mill Creek 1 | 55.0 | 60.4 | - | 4.26 |
| Mill Creek 3 | 42.6 | 66.0 | 1.25 | 4.03 |
| Paddy's Run 6 | 61.7 | 133.0 | - | 12.25 |
| Pleasants 1 | 119.1 | 152.9 | 1.46 | 7.94 |
| Pleasants 2 | 119.1 | 152.9 | 1.46 | 7.94 |
| Bruce Mansfield 1 | 120.7 | 144.2 | 3.28 | 11.31 |
| Bruce Mansfield 2 | 120.6 | 144.2 | 3.28 | 11.31 |
| Bruce Mansfield 3 | 213.6 | 181.1 | 3.11 | 10.27 |
| Hunter 1* | 56.7 | 75.1 | - | 4.66 |
| Hunter 2* | 60.2 | 72.6 | 0.32 | 4.55 |
| Huntington 1* | 63.0 | 90.0 | 1.27 | 5.89 |
| Average | 81.8 | 116.5 | 3.20 | 8.08 |
| Std. Dev. | 45.3 | 45.3 | 2.86 | 3.68 |

*Sulfur content below 1.5 percent.

Source: Reference 2.

KWh in capital and annual costs. This closeness in cost concurs with the results of the economic study by TVA in Table 28, showing that for low sulfur applications, lime is fairly competitive with limestone.

If transportation were a large fraction of the reagent cost, lime would have a greater advantage because of its lower molecular weight and subsequent greater reactivity per pound.

The 13 high sulfur lime systems average $141 per KW and 10.5 mills per KWh as opposed to the high sulfur limestone costs of $149 per KW and 9.2 mills per KWh. The trend shown in the cost studies of lower capital costs but higher annual costs is again confirmed by the commercial systems.

## Spray Drying

Spray drying FGD systems, because of several significant operational differences from wet systems, have generated a large amount of utility interest and acceptance for a new FGD technology. Both technical and economic advantages are responsible for the number of applications of this process. Table 30 lists the regulatory and operational status of spray drying systems. As can be seen, 12 of the 18 systems are required to meet or exceed the 1979 NSPS. All of these systems burn low sulfur coal (1.5 percent or less). As discussed at the beginning of this section, the variable $SO_2$ removal rate of the 1979 NSPS has allowed this technology to be used in many more applications than if a constant 90 percent removal limit had been promulgated.

### Process Description—

In spray drying systems, hot flue gas (typically ~300°F) is cooled by evaporation of a liquid or slurry to 20-50°F above the adibatic saturation temperature. The $SO_2$ is absorbed by the evaporating liquid and converted to a solid salt in the drying step. Figure 8 is a diagram of a typical spray drying system. The solid reaction product usually has less than one percent free moisture and along with fly ash is removed from the gas by a down

TABLE 30. OPERATIONAL AND REGULATORY STATUS OF SPRAY DRYING FGD SYSTEMS AS OF DECEMBER 1981

| Applicable $SO_2$ standard | MW Capacity (Number of Units) | | | | |
|---|---|---|---|---|---|
| | Operating | In construction | Contracted | Planned | Total |
| 1979 NSPS | - | 541 (4) | 1,060 (2) | 476 (2) | 2,077 (8) |
| 1971 NSPS | 440 (1) | - | 740 (2) | - | 1,180 (3) |
| >1979 NSPS | - | 1,010 (2) | 1,089 (2) | - | 2,099 (4) |
| >1971 NSPS, <1979 NSPS | - | 887 (2) | - | - | 887 (2) |
| ≤1971 NSPS | 110 (1) | - | - | - | 110 (1) |
| TOTAL | 550 (2) | 2,438 (8) | 2,889 (6) | 476 (2) | 6,353 (18) |

Source: References 2 and 51.

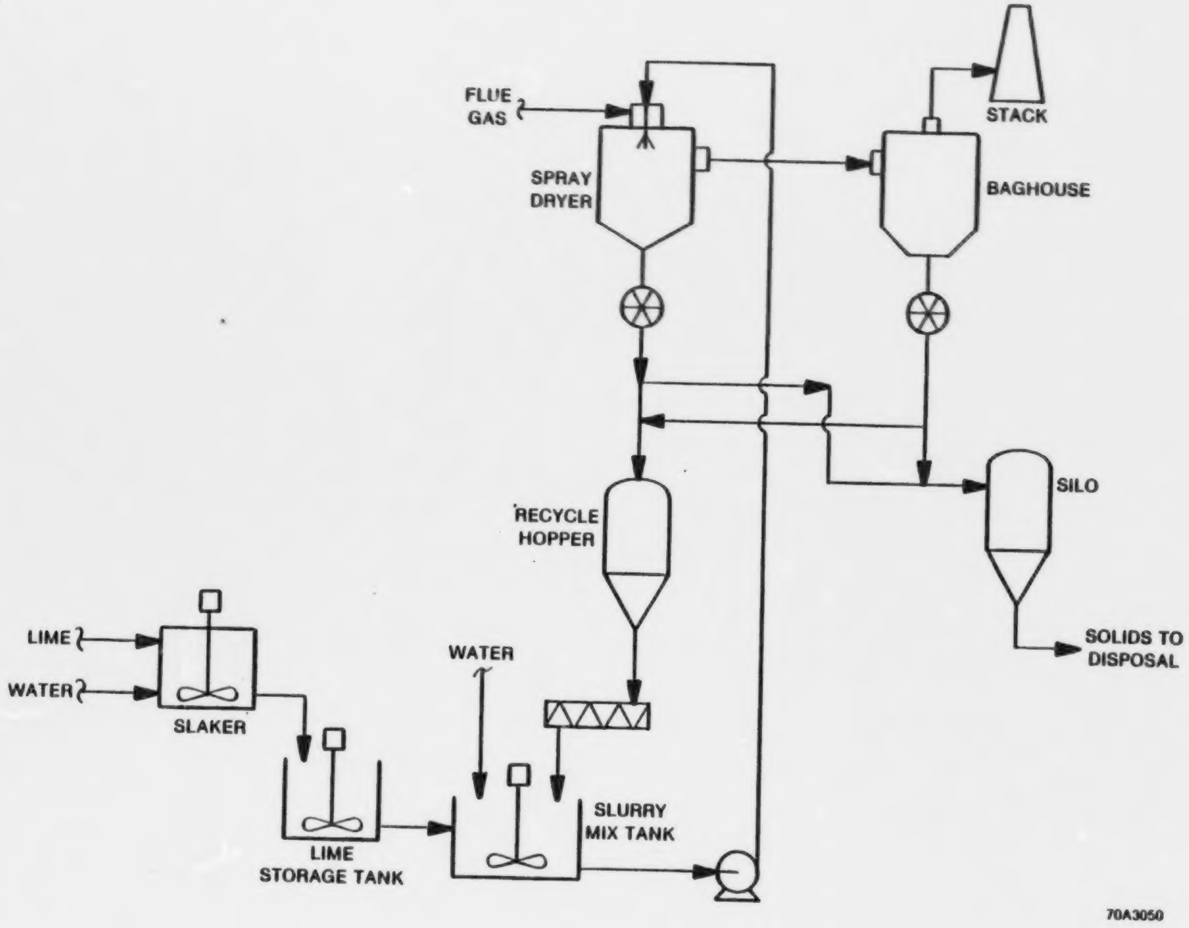

FLUE GAS
SPRAY DRYER
STACK
BAGHOUSE
SILO
RECYCLE HOPPER
SOLIDS TO DISPOSAL
LIME
WATER
SLAKER
WATER
LIME STORAGE TANK
SLURRY MIX TANK
70A3050

Figure 8. Spray Dryer Process Flow Diagram

stream ESP or baghouse. A portion of these solids are usually recycled in order to increase the reagent utilization. The majority of the planned systems use lime as the alkaline reagent. Sodium carbonate is also a satisfactory reagent. Many papers have been published recently further describing the principles of operation of spray drying systems (52, 53, 54, 55). The reaction products for lime systems are calcium sulfite and sulfate. Sodium sulfite and sulfate are formed when a sodium based system is used. The calcium based reactions are shown below.

$$SO_2(g) + Ca(OH)_2(l) \rightarrow CaSO_3(s) + H_2O(g)$$
$$SO_3(g) + Ca(OH)_2(l) \rightarrow CaSO_4(s) + H_2O(g)$$
$$1/2\ O_2(g) + CaSO_3(s) \rightarrow CaSO_4(s)$$

The reagent-flue gas contact time in the reactor is generally on the order of 10 seconds. During this period, the slurry liquor is evaporated, cooling the flue gas. The amount of liquid evaporating determines the exit flue gas temperature. This temperature is usually maintained at 20°F or higher above the adiabatic saturation temperature to prevent moisture condensation in the baghouse or other downstream equipment. The stoichiometric ratio (moles of reagent per mole of $SO_2$) is an important variable affecting the $SO_2$ removal. Ratio's of nearly 2 have been required to achieve 90 percent removal (55). In order to improve the reagent utilization and to reduce the reagent costs, a portion of the collected solids is often recycled back to the reactor. This increases the reagent utilization while maintaining a higher stoichiometry in the reactor. Some $SO_2$ removal across the baghouse filter cake has also been documented (56).

Disposal of spray dryer wastes usually requires addition of some amount of water for compaction and to prevent dusting (57). Lined landfill areas are used in most utility applications to prevent ground water contamination. Systems using sodium based reagents may require more sophisticated disposal techniques due to the high solubility of most sodium compounds.

### Commercial Designs—

Table 31 presents several design features of commercial spray drying systems. Only two systems have been commercially operated, the Riverside 6-7 and Coyote 1 units (58, 59). The Riverside system is the only retrofit unit and has been used for testing purposes. The Coyote 1 unit burns high sodium ash lignite and uses sodium carbonate as the alkali reagent. Consequently, the disposal of the soluble fly ash and FGD waste requires the same special practices. In other installations, lime is the preferred reagent because of lower costs, and the low solubility of calcium compounds which presents fewer disposal problems.

The only other system to use sodium carbonate as the alkali reagent is the Jim Bridger 2A system. This system is a test module which uses lime and sodium carbonate. Several commerical lime systems are presently in the start-up phase. Table 32 presents additional design features of the commercial spray drying systems. The typical system uses rotary atomizers at an approach to saturation of about 20°F and a baghouse for particulate control. When a close approach temperature is used, bypass of hot flue gas is often employed to prevent downstream moisture condensation. The bypass gas is obtained either before or after the air preheater. Hot gas taken before the air preheater increases the plant heat rate.

There are several design differences not shown in the tables which have been discussed in a recent EPRI report (60). Joy-Niro and General Electric use only one atomizer per module while the other vendors typically use three. The average module size is about 125 MW although Rawhide 1 and Holcomb 1 have 140 and 158 MW modules, respectively. The horizontal modules used by Babcock and Wilcox at Craig 3 and Laramie River 3 are 149 and 192 MW each. A typical 500 MW plant with a spare absorber would have five 125 MW cylindrical modules or four 167 MW horizontal units. There is an economic tradeoff between the higher total capacity of the horizontal system and the extra ductwork and piping of the cylindrical units.

TABLE 31. SPRAY DRYING FGD SYSTEM DESIGN

| Plant | Startup | Regulatory Status[a] | Percent Sulfur | Reagent | Absorber Removal (%) | Unit Removal (%) | Vendor[b] |
|---|---|---|---|---|---|---|---|
| Oxbow 1 | 00/85 | A | 0.6 | Lime | – | – | JN |
| Shiras 3 | 09/82 | A | 1.5 | Lime | – | 80 | GE |
| Metro Coal 1 | 00/90 | A | 1.0 | Lime | – | – | – |
| Jim Bridger 2A | 01/82 | A | 0.6 | Lime or Sod. Carb. | – | – | FL |
| | | | | | | | |
| Holcomb 1 | 09/83 | A | 0.5 | Lime | 86 | 86 | JN |
| Stanton 10 | 06/82 | A | 0.8 | Lime | – | 91 | RC |
| Grand River 2 | 03/85 | A | 0.9 | Lime | 85 | 85 | FL |
| Valmy 2 | 08/84 | A | – | Lime | – | – | – |
| | | | | | | | |
| Coyote 1 | 04/81 | B | 0.9 | Sod. Carb. | 70 | 70 | RIWF |
| Springerville 1 | 02/85 | B | 0.6 | Lime | – | 61 | JN |
| Springerville 2 | 03/87 | B | 0.6 | Lime | – | 61 | JN |
| | | | | | | | |
| Antelope Valley 2 | 10/85 | C | 0.7 | Lime | 89 | 89 | JN |
| Laramie River 3 | 11/82 | C | 0.5 | Lime | – | 85 | BW |
| Sherburne 3 | 00/85 | C | 1.0 | Lime | – | – | – |
| Rawhide 1 | 12/83 | C | 0.3 | Lime | – | 80 | JN |
| | | | | | | | |
| Antelope Valley 1 | 10/83 | D | 0.7 | Lime | – | 78 | JN |
| Craig 3 | 04/83 | D | 0.5 | Lime | 87 | 87 | BW |
| | | | | | | | |
| Riverside 6-7 | 11/80 | E | 1.2 | Lime | 90 | – | JN |

[a] Regulatory Status: A-1979 NSPS; B- 1971 NSPS; C - Standard more stringent than 1979 NSPS;
D - Standard more stringent than 1971 NSPS, less than 1979 NSPS;
E - Standard equal to or less than 1971 NSPS.

[b] Vendors: BW - Babcock and Wilcox; FL - Flakt; GE - General Electric; JN - Joy-Niro;
RC - Research Cottrell; RIWF - Rockwell International - Wheelabrator Frye.

Source: References 2 and 51.

## TABLE 32. ADDITIONAL SPRAY DRYER DESIGN FEATURES

| Plant | Atomizer type | Approach to saturation (°F) | Particulate Collection device | Gas Bypass (%)[a] |
|---|---|---|---|---|
| Oxbow 1 | Rotary | - | - | - |
| Shiras 3 | Rotary | 25 | Baghouse | 3.0-H |
| Jim Bridger 2A | Nozzle | 22 | ESP | 0 |
| Holcomb 1 | Rotary | 50 | Baghouse | 0 |
| Stanton 10 | Rotary | 20 | Baghouse | - |
| Grand River 2 | Rotary | - | ESP | - |
| Coyote 1 | Rotary | 50 | Baghouse | 0 |
| Springerville 1 | Rotary | 20 | Baghouse | 17.5-H |
| Springerville 2 | Rotary | 20 | Baghouse | 17.5-H |
| Antelope Valley 2 | Rotary | 20 | Baghouse | 0 |
| Laramine River 3 | Nozzle | - | ESP | 5.0-C |
| Rawhide 1 | Rotary | - | - | - |
| Antelope Valley 1 | Rotary | 20 | Baghouse | 3.2-H |
| Craig 3 | Nozzle | 25 | Baghouse | 0 |
| Riverside 6-7 | Rotary | Variable | Baghouse and ESP | 0 |

[a] - hot side before air preheater; C - cold side after air preheater.

Source: References 51 and 60.

Another design feature incorporated in most of the commercial systems is the use of solids recycle. In a spray dryer, the reagent slurry is in contact with the flue gas for only a few seconds. In order to obtain a high $SO_2$ removal efficiency, it is necessary to have a fairly high stoichiometric ratio, which can result in an excess of unreacted reagent. By recycling a portion of the collected solids, the reagent utilization can be significantly increased. Recycling also allows any unreacted alkaline fly ash to also remove additional $SO_2$.

### Recent Innovations—

Spray drying itself is a recent innovation as far as FGD technology is concerned, although it has been used for many years in other industries to produce dry solids. The two commercial systems have barely two years of operating experience between them. However, a large amount of testing has been performed by the vendors to substantiate their commerical designs. This experience is documented in several referenced sources and the interested reader is referred to them (51,60).

Further research is being conducted to expand the applicability of spray drying to high sulfur coals. Initially, it was thought that high sulfur coals could not economicaly meet the NSPS using spray drying due to high excess reagent requirements. Recent system guarantees are pushing the removal efficiency and the design sulfur content higher. Consequently, vendor research is focusing on high sulfur applications and reactor optimization.

### Economics—

The major reason for the large number of spray dryer applications in the recent past is economic. These systems are capable of meeting both $SO_2$ and particulate emissions at a lower cost than wet scrubbing systems. Both case studies and actual plant costs confirm this statement, although actual plant and operating cost data is limited at the present time.

Economic Studies-- Two separate studies of spray drying are detailed in Table 33, both of which were performed by TVA. Estimates 1-4 were sponsored by EPRI while 5-9 were done for the EPA. It is important to note that the design of the high sulfur coal systems are based on pilot data, not commercial systems.

Estimates 1-4 were made using TVA's old design requirements for the 1971 NSPS using a high sulfur coal. It has not been proven that spray dryers can meet the NSPS under these conditions. Cases 1 and 3 were made using TVA's economic premises while 2 and 4 used EPRI's economic bases. For either basis, the soda ash systems are more expensive than the lime systems. This is due to the higher transportation and reagent cost of soda ash to the hypothetical midwest plant location. Electric Power Research Institute capital costs are 10-15 percent higher than TVA's, due to higher indirect costs. The design basis was the same for both estimates. Higher process and project contingency charges, engineering and home office fees, and land costs, and different lifetime operating profiles in the EPRI bases account for the difference in the estimates.

Estimates 5-9, again by TVA, are more comparable to actual installations. Four different coal types are used; lignite, subbituminous, and high and low sulfur bituminous. The 1979 NSPS is used as a design basis, requiring higher $SO_2$ removal and a spare module. The costs have also been updated to mid-1982 for capital and 1984 for operating charges. All of the estimates include baghouses for particulate control. The design of each system reflects process optimization due to the fuel source. For example, solids recycle is used for Cases 5 and 7, which have an alkaline ash. Case 9 also uses recycle to reduce the lime stoichiometry. Case 6 uses soda ash which is sufficiently reactive, not requiring recycle. It is interesting to note that the cost differential between the soda ash and lime systems (Cases 6 and 7) is much less than either of the previous estimates, 1 and 3 (TVA economics) or 2 and 4 (EPRI economics). The major reason for this difference is the location of the power plant. Estimates 5, 6, and 7 are for a

TABLE 33. ESTIMATES OF NEW SPRAY DRYING FGD SYSTEMS

| Estimate Number | 1 | 2 | 3 | 4 |
|---|---|---|---|---|
| Source | TVA | TVA | TVA | TVA |
| Sponsor | EPRI | EPRI | EPRI | EPRI |
| Reference | 15 | 15 | 15 | 15 |
| Plant Size (MW) | 500 | 500 | 500 | 500 |
| Coal Type | Eastern | Eastern | Eastern | Eastern |
| Coal Sulfur (%) | 3.5 | 3.5 | 3.5 | 3.5 |
| NSPS | 1971 | 1971 | 1971 | 1971 |
| Base Year - Capital | 1979 | 1979 | 1979 | 1979 |
| Base Year - Operations | 1980 | 1980 | 1980 | 1980 |
| Reagent | Soda Ash | Soda Ash | Lime | Lime |
| Solids Recycle | No | No | No | No |
| Spare Module | No | No | No | No |
| Unit Removal (%) | 90 | 90 | 90 | 90 |
| Capital Cost ($/KW) | 106 | 121 | 90 | 102 |
| Revenue Requirements (mills/KWh) | | | | |
| First Year | 7.83 | 7.72 | 5.10 | 5.47 |
| Levelized | 9.07 | 11.43 | 6.15 | 7.68 |

(Continued)

TABLE 33. ESTIMATES OF NEW SPRAY DRYING FGD SYSTEMS (Continued)

| Estimate Number | 5 | 6 | 7 | 8 | 9 |
|---|---|---|---|---|---|
| Source | TVA | TVA | TVA | TVA | TVA |
| Sponsor | EPA | EPA | EPA | EPA | EPA |
| Reference | 14 | 14 | 14 | 14 | 14 |
| Plant Size (MW) | 500 | 500 | 500 | 500 | 500 |
| Coal Type | Lignite | Subbituminous | Subbituminous | Bituminous | Bituminous |
| Coal Sulfur (%) | 0.9 | 0.7 | 0.7 | 0.7 | 3.5 |
| NSPS | 1979 | 1979 | 1979 | 1979 | 1979 |
| Base Year - Capital | 1982 | 1982 | 1982 | 1982 | 1982 |
| Base Year - Operations | 1984 | 1984 | 1984 | 1984 | 1984 |
| Reagent | Lime | Soda Ash | Lime | Lime | Lime |
| Solids Recycle | Yes | No | Yes | No | Yes |
| Spare Module | Yes | Yes | Yes | Yes | Yes |
| Unit Removal (%) | 70 | 70 | 70 | 70 | 89 |
| Capital Cost ($/KW) | 165 | 159 | 154 | 151 | 200 |
| Revenue Requirements (mills/KWh) | | | | | |
| First Year | 7.61 | 7.42 | 6.92 | 6.76 | 11.60 |
| Levelized | 10.43 | 10.23 | 9.39 | 9.18 | 17.13 |

plant located in the Upper Great Plains and Rocky Mountain location. Since most of the soda ash deposits are in this region, reagent transportation costs are much less.

In order to fully compare the costs of spray drying systems, the second TVA study also costed limestone systems with ESPs for particulate control. This comparison is shown in Table 34 for 500 MW plants. As can be seen, lime spray drying has lower levelized costs than limestone FGD with ESPs for all three of the low sulfur coals. For the high sulfur coal, the levelized cost of the conventional limestone system is slightly less than the spray dryer. However, two factors should be noted. First, the design of the high sulfur spray dryer has not been commercially proven whereas limestone systems have an extensive operating history. Secondly, the limestone system in this estimate is the same as Estimate 5 in Table 17. Without the ESP and associated costs, the limestone unit alone costs $206 per KW with levelized charges of 15.08 mills per KWh. For the same application, an adipic acid enhanced system (Estimate 6, Table 17) was shown to be $20 per KW and 1.77 mills per KWh less than the limestone system. A comparison of an adipic acid system would show a substantial economic advantage over spray drying in this application besides being technically demonstrated.

Based on the results of these studies, it would appear that spray drying is economically competitive with limestone systems. As the coal sulfur decreases, spray drying becomes more attractive because of lower capital charges. Conversely, at high sulfur levels, the reagent cost differential between lime and limestone appears to become the controlling economic factor.

Commercial Spray Drying Systems Costs— No operating costs of commercial spray drying systems have been published yet, due to the newness of the technology. Capital costs for some of the systems are presented in Table 35. It should be emphasized that the system costs reported are often on different contract bases. Where the project scope is known, this information is identified.

TABLE 34. COMPARISON OF LIME SPRAY DRYER AND LIMESTONE FGD AND ESP

| Coal Type | Lignite | Subbituminous | Low sulfur bituminous | High sulfur bituminous |
|---|---|---|---|---|
| Capital cost ($/KW) | | | | |
| Lime spray dyrer | 165 | 154 | 151 | 200 |
| Limestone FGD and ESP | 215 | 176 | 185 | 244 (224)[a] |
| Revenue requirements (mills/KWh) | | | | |
| First year | | | | |
| Lime spray dryer | 7.61 | 6.92 | 6.76 | 11.60 |
| Limestone FGD and ESP | 9.57 | 7.90 | 8.34 | 11.78 (10.58)[a] |
| Levelized | | | | |
| Lime spray dryer | 10.43 | 9.39 | 9.18 | 17.13 |
| Limestone FGD and ESP | 12.96 | 10.73 | 11.35 | 16.43 (14.66)[a] |

[a] Adipic Acid Enhanced Limestone System and ESP.

Source: Reference 14

## TABLE 35. CAPITAL COSTS OF COMMERCIAL SPRAY DRYING SYSTEMS

| Unit | Capital Cost ($/KW) | Contract scope Included | Contract scope Excluded |
|---|---|---|---|
| Shiras 3 | 154 | Supply, erection, I.D. Fan | Foundations |
| Holcomb 1 | 189 | Supply, erection waste disposal | None |
| Stanton 10 | 95 | Supply, erection | Foundations, support steel |
| Coyote 1 | 84 | Supply, erection start up | Foundation, ash handling system |
| Springerville 1,2 | 60 each | Supply only | Foundations, erection |
| Antelope Valley 1 | 114 | Supply, erection, foundations | Lime storage |
| Antelope Valley 2 | 123 | Supply, erection, foundations | Lime storage |
| Laramine River 3 | 63 | Supply, erection | Foundations |
| Rawhide 1 | 70 | Supply, erection | Not reported |
| Craig 3 | 100 | Supply, erection | Not reported |

Source: Reference 60.

Table 36 presents estimated total costs of wet and dry FGD systems at the Laramie and Antelope Stations of Basin Electric. The dry systems are predicted to save 23 and 47 million dollars over the 35 year life of each plant in 1981 dollars (61). These savings are site specific, reflecting the available alkalinity of the fly ash, reagent transportation costs, and $SO_2$ removal requirements.

## Alkaline Ash

Alkaline ash FGD systems use the alkaline compounds present in fly ash (calcium, magnesium, and sodium oxides) to react with and remove $SO_2$. These systems are only used at plants burning coals which contain alkaline ash. These coals are typically low sulfur subbituminous or lignite fuels from Montana and Wyoming, although other states also have coal with alkaline ash.

Table 37 lists the operational and regulatory status of the alkaline ash FGD systems. As can be seen, this technology is not being planned for any new units. The new units which could conceivably use alkaline ash scrubbing are now using either limestone or spray drying systems.

Although the economics of alkaline ash scrubbing are often very favorable (since reduced quantities of reagents are required), there are several technical reservations about the process which will likely limit further applications. These considerations are discussed further below.

### Process Description—

Alkaline ash systems operate chemically in much the same manner as lime or limestone systems. In the absorber or reaction tank, calcium oxide is dissolved from the ash, hydrates, and neutralizes $SO_2$. Typically, alkaline ashes contain 10-20 percent soluble alkalis (CaO, MgO, $Na_2O$). The alkalis can be present in several different mineral forms, some of which are not soluble at the operating system pH. Forty to seventy percent of the alkalis are usually available for $SO_2$ neutralization (62). There are no accurate lab tests which can be used to determine how much alkalinity can be obtained

TABLE 36. BASIN ELECTRIC ECONOMIC EVALUATION

| Cost factor (1981 dollars) | Antelope Valley 1 |  | Laramine River 3 |  |
|---|---|---|---|---|
|  | Dry scrubber | Wet scrubber | Dry scrubber | Wet scrubber |
| Capital costs | $ 49,665,100 | $ 55,927,400 | $ 49,807,000 | $ 60,632,100 |
| Reagent cost | 38,587,100 | 42,505,000 | 48,880,000 | 9,367,000 |
| Power cost | 7,466,700 | 13,040,000 | 4,806,000 | 19,610,000 |
| Manpower cost | 15,925,000 | 36,750,000 | 15,925,000 | 36,750,000 |
| Replacement materials | 17,500,000 | 28,000,000 | 15,750,000 | 28,000,000 |
| Pressure loss | – | – | 4,623,800 | 8,678,600 |
| TOTAL | $129,144,300 | $176,224,400 | $139,791,800 | $163,037,700 |

Source: Reference 61.

TABLE 37. OPERATIONAL AND REGULATORY STATUS OF ALKALINE ASH FGD SYSTEMS AS OF DECEMBER 1981 (MW Scrubbed and Number of Units)

| Applicable $SO_2$ standard | Operating | In construction | Total |
|---|---|---|---|
| 1979 NSPS | 475 (1) | - | 475 (1) |
| 1971 NSPS | 1,374 (4) | - | 1,374 (4) |
| >1979 NSPS | 579 (3) | 1,400 (2) | 1,979 (5) |
| >1971 NSPS, <1979 NSPS | 1,665 (3) | - | 1,665 (3) |
| TOTAL | 4,093 (11) | 1,400 (2) | 5,493 (13) |

Source: Reference 2.

from the fly ash in a commercial FGD system. Consequently, it is necessary to run pilot tests using the ash in a plant environment. If insufficient amounts of alkalinity are available from the ash to achieve the desired removal, a supplemental reagent, either lime or limestone, can be used to increase the removal efficiency.

Because of the low sulfur levels of most alkaline ash coals, the oxidation level is usually very high, resulting in gypsum formation. This is due to the high oxygen to $SO_2$ ratio of the flue gas. One factor complicating the design of alkaline ash systems is the inverse alkalinity solubility-$SO_2$ removal relationship. As the liquor pH decreases, more alkalis can be dissolved from the ash. However, low pH levels reduce the $SO_2$ removal. Consequently, each system has some equilibrium pH-removal value. If the removal is too low and lime is added to raise the pH, the amount of alkalinity being supplied by the fly ash decreases. The lime addition rate can vary significantly depending on the ash composition and volume.

Commercial Designs—

Tables 38 and 39 present selected design features of the alkaline ash systems. All of the systems use low sulfur coal. Three of the plants (Coal Creek 1, 2 and M.R. Young 2) use hot-side ESPs to remove particulate matter. Alkaline ash is then fed to the absorbers as needed for pH control. In the other systems, a venturi is used for particulate control and $SO_2$ removal. Limestone is used as a supplemental additive only at Sherburne 1 and 2. All of the other units have lime facilities, although some systems do not use lime in normal operations.

Only the Four Corners units are retrofit installations. These venturis were initially used just for particulate control. Based upon a consent decree, APS added a lime feeding system to enhance the $SO_2$ removal. Other particulate-only systems scrubbling alkaline ash coals are shown in Table 40, along with typical $SO_2$ removal efficiencies. These units are not required to meet any $SO_2$ emission standards.

## TABLE 38. ALKALINE ASH FGD SYSTEMS

| Plant | Startup | Regulatory Status[a] | Percent Sulfur | L/G (gal/1000 acf) | Absorber Removal (%) | Unit Removal (%) | Absorber Type[b] | Vendor[c] |
|---|---|---|---|---|---|---|---|---|
| Clay Boswell 4 | 04/80 | A | 0.9 | 72[d] | 89 | 85 | V-ST | P |
| Coal Creek 1 | 07/79 | B | 0.6 | 79 | 90 | 54 | ST | CE |
| Coal Creek 2 | 07/80 | B | 0.6 | 79 | 90 | 54 | ST | CE |
| Colstrip 1 | 09/75 | B | 0.8 | 37[d] | 60 | 60 | V-T | THY |
| Colstrip 2 | 05/76 | B | 0.8 | 37[d] | 60 | 60 | V-T | THY |
| Four Corners 1 | 11/79 | C | 0.8 | 9 | 80 | 80 | V | GE |
| Four Corners 2 | 11/79 | C | 0.8 | 9 | 80 | 80 | V | GE |
| Four Corners 3 | 11/79 | C | 0.8 | 9 | 80 | 80 | V | GE |
| Colstrip 3 | 10/83 | C | 0.7 | - | - | - | - | THY |
| Colstrip 4 | 00/84 | C | 0.7 | - | - | - | - | THY |
| M.R. Young 2 | 09/77 | D | 0.7 | 72 | 85 | 36 | ST | THY |
| Sherburne 1 | 03/76 | D | 0.8 | 31[d] | 50 | 50 | V-ST | CE |
| Sherburne 2 | 03/77 | D | 0.8 | 31[d] | 50 | 50 | V-ST | CE |

[a] Regulatory Status: A - 1979 NSPS; B- 1971 NSPS; C - Standard more stringent than 1979 NSPS;
D - Standard more stringent than 1971 NSPS, less than 1979 NSPS.

[b] Absorber Type: ST - spraytower; T - tray tower; V - venturi.

[c] Vendors: CE - Combustion Engineering; GE - General Electric; P - Peabody; THY - Thyssen/CEA.

[d] Includes venturi L/G.

Source: Reference 2.

TABLE 39. OTHER DESIGN FEATURES - ALKALINE ASH SYSTEMS

| | Particulate Control | | |
|---|---|---|---|
| Plant | Application | Device | Additive |
| Clay Boswell 4 | New | Venturi | Lime |
| Coal Creek 1, 2 | New | ESP | Lime |
| Colstrip 1, 2 | New | Venturi | Lime |
| Four Corners 1, 2, 3 | Retrofit | Venturi | Lime |
| M.R. Young 2 | New | ESP | Lime |
| Sherburne 1, 2 | New | Venturi | Limestone |

Source: Reference 2.

TABLE 40. OTHER ALKALINE ASH PARTICULATE SCRUBBERS

| Plant-Utility | Contractor | Sulfur (%) | $SO_2$ Removal (%) |
|---|---|---|---|
| Dave Johnston 4-PP&L | Venturi | 0.5 | 40 |
| Arapahoe 4-PSC | TCA | 0.7 | 20 |
| Cherokee 1, 2, 4-PSC | TCA | 0.7 | 50 |
| Valmont 5-PSC | TCA | 0.7 | 50 |
| Aurora 1, 2-MP&L | Spray impingement | 0.9 | 37 |
| Clay Boswell 1-MP&L | Spray impingement | 0.9 | 37 |

Source: Reference 63.

The method selected for particulate removal, wet scrubbing or dry removal, depends on the utility's preferences. Wet scrubbing units using venturis and TCAs may not satisfy the particulate requirements of the 1979 NSPS (5). Current limits are 0.03 pound of particulates per million Btu's of heat input. Although it has been shown that ESPs and baghouses can meet this level; wet collection devices do not always achieve it. Additionally, in a wet system, all of the fly ash is disposed of as a wet sludge. This eliminates the possibility of using the ash in road aggregate or as some other by-product. When the fly ash is collected upstream of the scrubbing system, only the amount required for $SO_2$ removal is used. For a 0.8 percent sulfur, 15 percent ash (5 percent available alkalinity), 8000 Btu per pound coal, only one third of the ash is required to neutralize the $SO_2$. This option increases the capital cost of the system since separate particulate and $SO_2$ control systems are needed.

Table 41 presents a list of other units which use combined particulate and $SO_2$ control. However, the ash at these units is not alkaline, therefore, these systems are classified according to their respective FGD processes.

### Recent Innovations—

All of the alkaline ash systems in Table 38 were initially tested at the pilot plant level to ensure that adequate $SO_2$ removal could be obtained (63). At the present time, however, no other utilities are considering this type of system.

Some research has been performed using humic acids to increase the soluble alkali fraction present in fly ash. Humic acids are complex organic polymers of uncertain structure which contain carboxylic and phenolic groups which may be extracted from coal, peat, and soil (64). These acids perform in much the same manner as adipic acid and limestone. The humic acid helps solubilize not only the basic oxides in the fly ash, but also some of the aluminum and iron oxides (65). As of yet, these tests have only been conducted at the bench-scale level.

### TABLE 41. NON-ALKALINE ASH COMBINED PARTICULATE - $SO_2$ CONTROL SYSTEMS

| Unit | FGD Process |
|---|---|
| Cholla 1, 2 | Limestone |
| LaCygne 1 | Limestone |
| Winyah 2 | Limestone |
| Lawrence 5 | Limestone |
| Widows Creek 7, 8 | Limestone |
| Elrama 1-4 | Lime |
| Phillips 1-6 | Lime |
| Green River 1-3 | Lime |
| Bruce Mansfield 1, 2 | Lime |
| Reid Gardner 1, 2, 3 | Sodium carbonate |
| Delaware City 1, 2, 3 | Wellman-Lord |

Source: Reference 2

Economics--

Only one economic study using alkaline ash was found in the literature (66). The results of this study are shown in Table 42. This system was estimated to have a capital cost 2-3 percent higher than a comparable lime system because of the fly ash silos, larger pumps, and higher L/G. Annual costs were several percent lower. It should be noted that no particulate collection costs were included in this estimate. Comparing an ESP-lime system to venturi-alkaline ash unit would probably show a significant capital cost reduction for the combined system.

Actual and adjusted costs of alkaline ash systems for commercial units are shown in Table 43. The costs shown in this table are only for the $SO_2$ removal portion of the process. In discussions with each utility, PEDCo adjusts the total system cost to only show the $SO_2$ control cost. The average adjusted cost of $118 per KW is much higher than the low sulfur lime capital cost, $73 per KW. Differences in these costs are most likely due to site specific variables; design conditions, spare modules, materials of contruction, and other factors.

## Sodium Carbonate

Sodium carbonate FGD systems are limited to applications in the Western U.S. All of the systems, planned and operating, are located in dry regions where the liquid waste can be crystallized in evaporation ponds. The plants are also located near the reagent source. The operational and regulatory status of these plants is presented in Table 44. Three of the four planned units are to be located at the Jim Bridger station of Pacific Power and Light. However, the utility has built a spray dryer test unit at this site to explore both lime and sodium carbonate usage. It is possible that these systems will eventually use spray drying rather than wet sodium carbonate scrubbing.

TABLE 42. ESTIMATE OF NEW ALKALINE ASH FGD SYSTEM

| Estimate Number | 1 |
|---|---|
| Source | ERDA |
| Sponsor | EPA |
| Reference | 66 |
| Plant size (MW) | 500 |
| Coal type | Lignite |
| Coal sulfur (%) | 0.8 |
| NSPS | 1971 |
| Base year - capital | 1976 |
| Base year - operations | 1976 |
| Absorber type | Spray tower |
| Absorber removal (%) | 85 |
| Spare module | No |
| L/G (gallons/1000 acf) | 80 |
| Capital Cost ($/KW) | 65 |
| Revenue Requirements (mills/KWh) | |
| First year | 1.93 |

## TABLE 43. OPERATIONAL ALKALINE ASH SYSTEM COSTS

| Unit | Capital costs ($/KW) Reported | Capital costs ($/KW) Adjusted | Annual Costs (mills/KWh) Reported | Annual Costs (mills/KWh) Actual |
|---|---|---|---|---|
| Coal Creek 1 | 43.4 | 52.5 | 0.41 | 2.99 |
| Coal Creek 2 | 43.4 | 52.5 | 0.41 | 2.99 |
| Clay Boswell 4 | 173.8 | 184.4 | 5.38 | 14.09 |
| M.R. Young 2 | 100.3 | 155.7 | 0.53 | 6.44 |
| Colstrip 1 | 101.4 | 145.9 | 2.97 | 8.27 |
| Sherburne 1 | 49.3 | 102.6 | 0.75 | 5.43 |
| Sherburne 2 | 49.3 | 102.6 | 0.75 | 5.43 |
| Average | 82.8 | 117.7 | 1.77 | 6.74 |
| Std. Dev. | 45.7 | 48.5 | 1.82 | 3.59 |

Source: Reference 2.

TABLE 44. OPERATIONAL AND REGULATORY STATUS OF SODIUM CARBONATE SYSTEMS AS OF DECEMBER 1981

| Applicable $SO_2$ Standard | MW (Number of Units) | | |
|---|---|---|---|
| | Operating | Planned | Total |
| 1979 NSPS | - | 1,900 (4) | 1,900 (4) |
| 1971 NSPS | 375 (3) | - | 375 (3) |
| >1979 NSPS | 330 (1) | - | 330 (1) |
| >1971 NSPS, ≤1979 NSPS | 550 (1) | - | 550 (1) |
| TOTAL | 1,255 (5) | 1,900 (4) | 3,155 (9) |

Source: Reference 2.

### Process Description—

The sodium carbonate process uses a clear liquor composed of sodium carbonate and sodium sulfite to remove $SO_2$. The high alkalinity of this system, which is directly related to the high solubility of the sodium reagent, allows for very efficient $SO_2$ removal at very low L/G ratios. The reactions are as shown below.

$$SO_2(g) + Na_2CO_3(l) \rightleftarrows Na_2SO_3(l) + CO_2(g)$$

$$SO_2(g) + H_2O + NaSO_3(l) \rightleftarrows 2\ Na(HSO_3)(l)$$

$$SO_3^{=} + 1/2\ O_2(g) + 2\ Na^{+} \rightarrow Na_2SO_4(l)$$

The spent liquor from the absorber is generally sent to a series of evaporation ponds. As the water evaporates, sodium sulfite crystallizes. This salt cake requires careful disposal to prevent groundwater contamination. There is a potential market for this salt cake in Kraft pulping and detergent production.

### Commercial Designs—

Table 45 presents design features of the commercial sodium carbonate systems. Due to the high removal capability of the liquor, a low L/G is adequate. Tray towers are used to provide sufficient contact surface area between the gas and liquid phases. The Reid Gardner units use both sodium carbonate and trona, a mineral form of sodium carbonate, bicarbonate, and chloride. At these units, a venturi is also used for particulte control. The Jim Bridger units are located near massive deposits of sodium carbonate at Green River, Wyoming. As mentioned previously, although three more facilities at the Bridger station are planned to use wet scrubbing with sodium carbonate, spray dryer testing at this location could result in a change of processes. In this event, the ranking of sodium carbonate scrubbing relative to the other FGD systems would drop considerably.

### TABLE 45. SODIUM CARBONATE FGD SYSTEMS

| Plant | Startup | Regulatory Status[a] | Percent Sulfur | L/G (gal/1000acf) | Absorber Removal(%) | Unit Removal(%) | Absorber Type[b] | Vendor[c] |
|---|---|---|---|---|---|---|---|---|
| Reid Gardner 4 | 4-83 | A | 0.8 | – | – | – | – | THY |
| Jim Bridger 1 | 0-88 | A | 0.6 | – | – | – | – | – |
| Jim Bridger 2 | 0-90 | A | 0.6 | – | – | – | – | – |
| Jim Bridger 3 | 0-86 | A | 0.6 | – | – | – | – | – |
| Reid Gardner 1 | 3-74 | B | 0.5 | 10[d] | 91 | 90 | V-T | THY |
| Reid Gardner 2 | 4-74 | B | 0.5 | 10[d] | 91 | 90 | V-T | THY |
| Reid Gardner 3 | 6-76 | B | 0.5 | 10[d] | 91 | 85 | V-T | THY |
| Naughton 3 | 9-81 | C | 0.6 | – | – | 70 | – | UOP |
| Jim Bridger 4 | 9-79 | D | 0.6 | 20 | 91 | 91 | T | UOP |

[a] Regulatory Class: A-1979 NSPS; B-1971 NSPS; C-Standard more stringent than 1979 NSPS; D-Standard more stringent than 1971 NSPS, less than 1979 NSPS.

[b] Absorber Type: T-tray tower; V-venturi

[c] Vendors: THY-Thyssen-CEA; UOP-Air Correction Division

[d] Includes venturi L/G

Source: Reference 2.

Economics--

Only one study was found which estimated the cost of a sodium carbonate systems. This estimate, for a 3.5 percent sulfur coal at a 500 MW plant had a capital cost of $65 per KW (67). A limestone system in the same study was estimated to cost $96 per KW in 1979 dollars for 90 percent $SO_2$ control. Operating costs for the sodium carbonate system were not estimated. They are, however, very dependent on the reagent transportation costs and the waste disposal procedure.

Reported and adjusted capital and operating costs for the commercial sodium carbonte systems are presented in Table 46. The average adjusted capital and operating costs are approximately equal to the cost of the limestone systems scrubbing low sulfur coal.

TABLE 46. OPERATIONAL SODIUM CARBONATE SYSTEM COSTS

| | Capital Costs ($/KW) | | Annual Costs (mills/KWh) | |
|---|---|---|---|---|
| Unit | Reported | Adjusted | Reported | Adjusted |
| Reid Gardner 1 | 42.9 | 87.1 | 0.46 | 5.78 |
| Reid Gardner 2 | 42.9 | 87.1 | 0.46 | 5.78 |
| Reid Gardner 3 | 100.8 | 150.9 | 0.23 | 7.40 |
| Jim Bridger 4 | 90.3 | 118.4 | - | 6.81 |
| Average | 69.2 | 110.9 | 0.38 | 6.44 |
| Std. Dev. | 30.7 | 30.5 | 0.13 | 0.80 |

Source: Reference 2

## Dual Alkali

Dual alkali systems, as the name implies, use two chemical reagents to remove $SO_2$ from flue gas and produce a throwaway sludge. In the most common arrangement, sodium carbonate is used to remove $SO_2$, forming soluble sodium sulfite, which is then reacted with lime to produce a solid waste. The sodium solution is then recycled back to the absorber. Many other types of dual alkali systems have been proposed; ammonia-lime, magnesium-lime, aluminum hydroxide-limestone (DOWA) and sodium carbonate-limestone (limestone dual alkali). Tests of these last two systems have been performed at the prototype level.

Only 6 utility systems are or will be using the dual alkali system in the near future. The status of these systems is presented in Table 47. In contrast to the small percentage of dual alkali systems in use on utility systems, 27 of the 38 industrial boilers not using once through sodium scrubbing use dual alkali systems (68). The 183 once through sodium systems are primarily used for thermally enhanced oil recovery operations in Kern County, California and are atypical installations. Dual alkali systems are favored by industrial users due to the high reliability associated with clear liquor scrubbing. In utility applications, due to the higher reagent cost of lime over limestone, dual alkali units are at a disadvantage.

Process Description—

The dual alkali process uses alkaline sodium compounds to achieve efficient $SO_2$ removal, and lime to regenerate the sodium-based scrubbing liquor (69). A typical flow diagram is shown in Figure 9. The flue gas is normally treated for particulate removal before it reaches the system. A presaturator can be included to prevent chloride buildup in the recirculating liquor when treating high chloride coals. The dual alkali process has two major chemistry steps, absorption and regeneration. Both sodium carbonate and caustic can be used as the makeup alkali. Sodium sulfite is the major alkaline species presented in the liquid phase. The various sorption reactions are shown below.

$$SO_2(1) + 2NaOH(1) \rightleftarrows Na_2SO_3(1) + H_2O$$
$$SO_2(1) + NaOH(1) \rightleftarrows NaHSO_3(1)$$
$$SO_2(1) + NaHCO_3(1) \rightleftarrows Na_2HSO_3(1) + CO_2(g)$$
$$SO_2(1) + NaHCO_3(1) \rightleftarrows NaHSO_3(1) + CO_2(g)$$
$$SO_2(1) + Na_2SO_3(1) \rightleftarrows 2NaHSO_3(1)$$

Oxidation of the sulfite ion to sulfate also occurs.

TABLE 47. OPERATIONAL AND REGULATORY STATUS OF DUAL ALKALI SYSTEMS AS OF DECEMBER 1981 (MW scrubbed and number of units)

| Applicable $SO_2$ Standard | Operating | In Construction | Contracted | Total |
|---|---|---|---|---|
| 1979 NSPS | - | 421 (1) | 686 (2) | 1,107 (2) |
| 1971 NSPS | 882 (2) | - | - | 882 (2) |
| >1971 NSPS, ≤1979 NSPS | 299 (1) | - | - | 299 (1) |
|  | 1,181 (3) | 421 (1) | 686 (2) | 2,288 (6) |

Source: Reference 2.

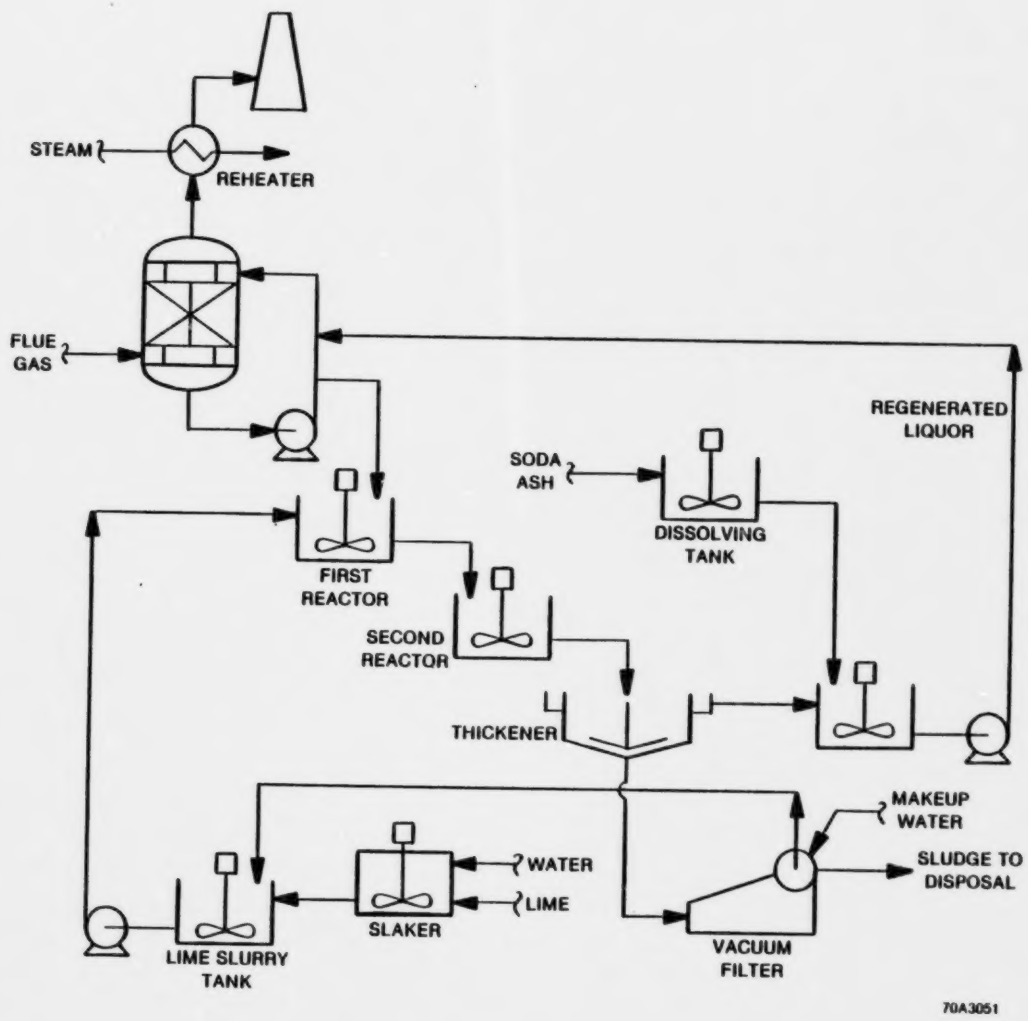

STEAM
REHEATER
FLUE GAS
REGENERATED LIQUOR
SODA ASH
DISSOLVING TANK
FIRST REACTOR
SECOND REACTOR
THICKENER
MAKEUP WATER
WATER
LIME
SLUDGE TO DISPOSAL
SLAKER
LIME SLURRY TANK
VACUUM FILTER
70A3051

Figure 9. Dual Alkali Process Flow Diagram

$$SO_3^{=} + 1/2\ O_2(g) \rightarrow SO_4^{=}$$

As seen in the above reactions caustic and soda ash form sodium sulfite, which upon additional sorption, forms sodium bisulfite. A portion of this recirculating clear liquor stream is sent to a reaction tank where lime is added to precipitate the sulfur species.

The reactions occurring in the regeneration section are given below.

$$Ca(OH)_2(l) + 2NaHSO_3(l) \rightleftarrows CaSO_3 \cdot 1/2\ H_2O(s) + Na_2SO_3(l) + 3/2H_2$$

$$Ca(OH)_2(l) + Na_2SO_3(l) + 1/2\ H_2O \rightleftarrows CaSO_3 \cdot 1/2\ H_2O(s) + 2NaOH(l)$$

$$Ca(OH)_2(l) + Na_2SO_4(l) + 2H_2O \rightleftarrows CaSO_4 \cdot 2H_2O(s) + 2NaOH(l)$$

The solids are separated in a thickener and concentrated in a vacuum filter before disposal. Limestone has also been tested as an alkali reagent for precipitation.

$$CaCO_3(l) + 2NaHSO_3(l) + 1/2\ H_2O \rightleftarrows CaSO_3 \cdot 1/2\ H_2O(s) + Na_2SO_3(l) + CO_2(g) + H_2O$$

The regenerated sodium compounds are then recycled to the absorber. Alkali losses due to the filter cake moisure are replaced by the makeup sodium reagent. Most systems include some type of filter cake washing system to minimize the sodium loss.

The oxidation reaction of sulfite to sulfate requires a treatment scheme to remove the inactive sulfate from the clear liquor. Both calcium sulfate and calcium sulfite are relatively insoluble compounds, although gypsum is about an order of magnitude more soluble than calcium sulfite. Consequently, unless the sulfate to sulfite ion ratio in the solution is the same as the ratio of solubility products of the two compounds, only one species should precipitate. In commercial operation and during test programs it has not been possible to produce gypsum crystals in a dual alkali liquor (70). The gypsum is removed from solution by a coprecipitation mechanism with the calcium sulfite. The sulfite and sulfate form a solid solution in the crystal lattice structure. This mechanism maintains the subsaturated gypsum condition without the presence of solid phase gypsum.

Commercial Designs—

Table 48 lists design parameters for the commercial dual alkali systems. The sulfur content of the units with dual alkali systems average much higher (3.4 percent) than any previous system discussed. Also, the L/G ratio is very low compared to lime and limestone systems. The absorbers use various internal contacting devices to permit efficient mass transfer at the low L/G ratios. Spray towers would require higher L/G levels to achieve good removal. The impingement tower used by FMC is of the disc and donut design although they also offer a spray tower design (70).

The Cane Run system of Louisville Gas and Electric was tested under the EPA dual alkali demonstration program. This program has tested the dual alkali process developed by Combustion Equipment Associates and Arthur D. Little at pilot, prototype, and commercial scale systems. The prototype testing was performed at Plant Scholz of Gulf Power Company over a 17 month period (71,72). Cane Run 6 is a 300 MW boiler owned by Louisville Gas and Electric. The EPA emission monitoring program at this site is discussed further in Section 3.

The Newton Station uses a high chloride coal (0.4 percent) and is designed for closed loop operation. A precooler is used to remove HCL from the flue gas to precent deactivation of the scrubbing solution. Purge liquor from the precooler is used in the final wash step of the filter cake to reduce sodium discharges from the process (73). This system was designed by Buell-Envirotech which is now a part of General Electric Environmental Services Company.

## TABLE 48. DUAL ALKALI FGD SYSTEM DESIGN

| Plant | Startup | Regulatory Status[a] | Percent Sulfur | L/G (gal/1000 acf) | Absorber Type[b] | Unit Removal (%) | Vendor[c] |
|---|---|---|---|---|---|---|---|
| A.B. Brown 2 | 1-85 | A | 3.4 | – | I | 90 | FMC |
| Schahfer 17 | 6-83 | A | 3.2 | – | – | – | FMC |
| Schahfer 18 | 6-85 | A | 3.2 | – | – | – | FMC |
| Newton 1 | 9-79 | B | 2.3 | 10 | P | 90 | GE |
| A. B. Brown 1 | 3-79 | B | 3.4 | 10 | I | 85 | FMC |
| Cane Run 6 | 4-79 | D | 4.8 | 10 | T | 95 | THY |

[a] Regulatory Class: A-1979 NSPS; B-1971 NSPS; D-Standard more stringent than 1971 NSPS, less than 1979 NSPS.

[b] Absorber Type: I-Impingement, P-Packed, T-Tray

[c] Vendors: FMC-FMC; GE-General Electric; THY-Thyssen-CEA

Source: References 2 and 70.

FMC, the third dual alkali vendor, has sold many industrial boiler FGD units, as well as the A.B. Brown system (74). With the exception of the "disc and donut" absorber tower, the process design is the same as that described previously (75,76). FMC is also building the Schahfer units for Northern Indiana Public Service. Very little design information has been published about these units. FMC has also recently developed a limestone dual alkali process which has lower annualized costs than the lime dual alkali process (70).

Recent Innovations—

The use of limestone rather than lime has been tested in the U.S. and is used commercially in Japan. Due to the slower dissolution rate of limestone, several reaction tanks in series, often heated, are required to obtain good limestone utilization. Tests at the prototype unit at Plant Scholz with limestone addition have recently been completed. The results were favorable (77). Laboratory work at RTP has also investigated the use of limestone as the regenerant (78). As stated previously, FMC is marketing a limestone dual alkali process (70). In Japan, 7 power plants totaling 2050 MW use limestone in a dual alkali system (79). Gypsum is produced in an acidification and oxidation step.

Economics—

Table 49 presents the estimated costs of several dual alkali systems. Estimates 1 and 2, by TVA are based on their old economic and design basis, including the 1971 NSPS. The plants are identical except for the coal sulfur content, 3.5 versus 2.0 percent. Both capital and operating costs are lower for the 2.0 percent sulfur case due to smaller material handling requirements. Estimates 3 and 4, by Bechtel, include spare modules and bypass ductwork. For both studies, the capital and operating costs are very similar to lime and limestone system costs estimated on the same basis.

TABLE 49. ESTIMATES OF NEW DUAL ALKALI FGD SYSTEMS

| Estimate Number | 1 | 2 | 3 | 4 |
|---|---|---|---|---|
| Source | TVA | TVA | Bechtel | Bechtel |
| Sponsor | EPA | EPA | EPRI | EPRI |
| Reference | 13 | 13 | 10 | 10 |
| Plant Size (MW) | 500 | 500 | 2 x 500 | 2 x 500 |
| Coal Type | Eastern | Eastern | Eastern | Western |
| Coal Sulfur (%) | 3.5 | 2.0 | 4.0 | 0.48 |
| NSPS | 1971 | 1971 | 1978[a] | 1978[a] |
| Base Year-Capital | 1979 | 1979 | 1978 | 1978 |
| Base Year-Operations | 1980 | 1980 | 1978 | 1978 |
| Absorber Type | Tray | Tray | Packed | Packed |
| Absorber Removal (%) | 79 | 63 | 85 | 83 |
| Spare Module | No | No | Yes | Yes |
| L/G (gallons/1000 acf) | 7 | 7 | 8 | 10 |
| Sludge Disposal | Pond | Pond | Landfill | Landfill |
| Capital Cost ($/KW) | 101 | 84 | 166 | 134 |
| Revenue Requirements (mills/KWh) | | | | |
| First Year | 4.2 | 3.2 | 9.3 | 6.1 |
| Levelized | 5.3 | 4.2 | 13.5 | 8.2 |

[a] Proposed 1978 NSPS

Actual and adjusted costs of the three commercial dual alkali systems are presented in Table 50. Again, the average capital and operating costs for these three systems is very close to the high sulfur lime and limestone commercial system costs presented in Tables 21 and 29. The Newton unit has a much higher capital cost than the other two systems due to expensive alloy materials and oversized solids handling equipment. This equipment is sized to handle the effluent from a second unit.

## Wellman-Lord

The Wellman-Lord FGD system is one of two regenerable systems in use on utility boilers. Regenerable systems produce elemental sulfur or sulfuric acid as a saleable by-product rather than a waste liquid or solid stream. Regenerable systems have not found widespread use in the utility industry for three reasons. First, they are usually more complicated than the throwaway systems due to the extra sulfur processing steps. Secondly, they are also more expensive to build and operate than the simpler systems. Finally, most utilities have the land area available to dispose of the sludge produced by throwaway systems. However, as land costs rise, especially in urban regions, more regenerable systems may be built.

The operational and regulatory status of the utility Wellman-Lord Systems is presented in Table 51. The four plants with the stringent $SO_2$ standard are located at the San Juan station of Public Service of New Mexico. Of the other four units, one is the Dean H. Mitchell retrofit system while the other three are on the Delaware City units of Delmarva Power Company. At all of these installations, special considerations unique to each site resulted in the selection of the Wellman-Lord system. The Wellman-Lord process is also used at over 30 industrial sites (boilers, Claus units, acid plants) in the U.S., Japan, and Germany (80).

### Process Description—

The Wellman-Lord process uses the reversible sorption and desorption reactions between sodium sulfite and $SO_2$ to produce a concentrated off-gas.

TABLE 50. OPERATIONAL DUAL ALKALI SYSTEM COSTS

| Unit | Capital Costs ($/KW) |  | Annual Costs (mills/KWh) |  |
|---|---|---|---|---|
|  | Reported | Adjusted | Reported | Adjusted |
| Newton 1 | 174.8 | 263.9 | - | 13.94 |
| Cane Run 6 | 71.5 | 87.8 | - | 7.23 |
| A. B. Brown 1 | 47.2 | 88.4 | 1.30 | 4.98 |
| Avg. | 97.8 | 146.7 | - | 8.72 |
| Std. Dev. | 67.8 | 101.5 | - | 4.66 |

Source: Reference 2.

TABLE 51. OPERATIONAL AND REGULATORY STATUS OF WELLMAN-LORD FGD SYSTEMS AS OF DECEMBER 1981 (MW scrubbed and number of units)

| Applicable SO Standard | Operating | In Construction | Total |
|---|---|---|---|
| >1979 NSPS | 1,245 (3) | 534 (1) | 1,779 (4) |
| ≤1971 NSPS | 295 (4) | - | 295 (4) |
| Total | 1,540 (7) | 534 (1) | 2,074 (8) |

Source: Reference 2.

A process flow diagram is shown in Figure 10. A prescrubber is used to remove chlorides and particulates before the gas is scrubbed. $SO_2$ reacts with sodium sulfite in the absorber as shown below.

$$SO_2(g) + Na_2SO_3(1) + H_2O \rightleftarrows 2NaHSO_3(1)$$

Oxidation of the sulfite ion also occurs, forming sodium sulfate.

$$Na_2SO_3(1) + 1/2\ O_2(g) \rightarrow Na_2SO_4(1)$$

Spent liquor from the absorber is filtered to remove any fly ash solids before being processed in the evaporators. In the evaporators, the absorption reaction is reversed, liberating $SO_2$ and water vapor.

$$2NaHSO_3(1) \rightleftarrows SO_2(g) + H_2O(g) + Na_2SO_3(s)$$

Sodium sulfite crystals are formed due to evaporative concentration of the purge liquor via the above reaction. The $SO_2$-rich gas is cooled to remove most of the water vapor prior to processing in a sulfuric acid or elemental sulfur plant. The sodium sulfite slurry is dissolved along with makeup sodium carbonate and returned to the absorber.

Unreactive sodium sulfite is removed from the liquor in a high temperature crystallization step. Earlier system designs used a more energy intensive low-temperature process which removed sodium sulfate from solution because of its lower solubility at lower temperatures. In the new treatment method, a portion of the absorber effluent is heated, driving off steam and $SO_2$, which concentrates the solution. Both sodium sulfate and sulfite crystals are formed, which are centrifuged and dried to a powder consistency. This salt cake is stored in silos and sold as as by-product to the paper industry.

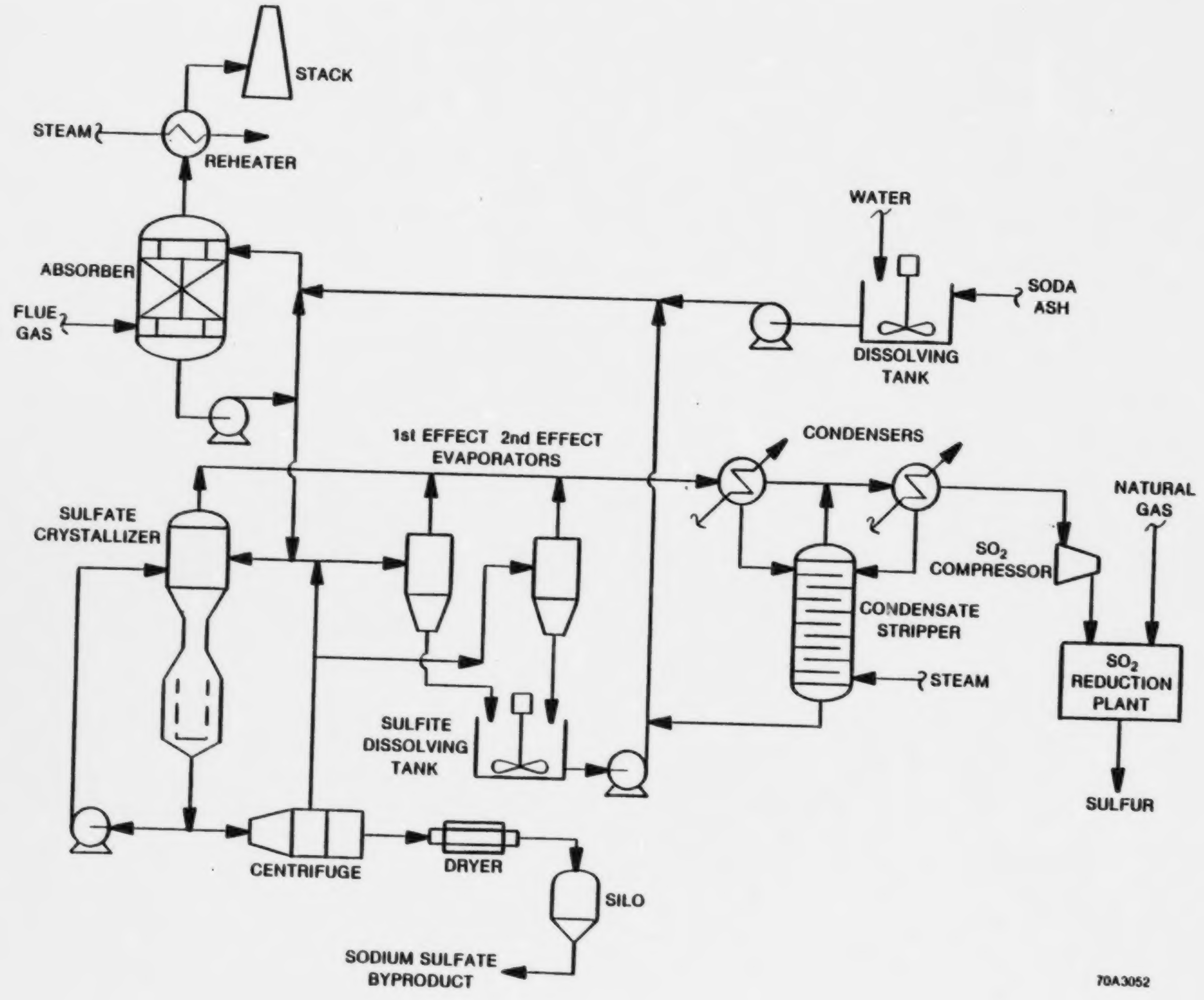

STACK
STEAM
REHEATER
ABSORBER
FLUE GAS
WATER
SODA ASH
DISSOLVING TANK
1st EFFECT 2nd EFFECT
EVAPORATORS
CONDENSERS
NATURAL GAS
SULFATE CRYSTALLIZER
$SO_2$ COMPRESSOR
CONDENSATE STRIPPER
STEAM
$SO_2$ REDUCTION PLANT
SULFITE DISSOLVING TANK
SULFUR
CENTRIFUGE
DRYER
SILO
SODIUM SULFATE BYPRODUCT
70A3052

Figure 10. Wellman-Lord Process Flow Diagram

The concentrated $SO_2$ steam can be converted into either sulfuric acid or sulfur. If acid is produced, the $SO_2$ must be diluted prior to entering the catalytic oxidation beds. Elemental sulfur is produced at three of the eight utility installations. The Allied $SO_2$ reduction system uses natural gas (a coal process has recently been developed) to produce sulfur. The hydrocarbon and part of the $SO_2$ stream are partially reduced in a catalytic reactor forming sulfur and $H_2S$. The sulfur is condensed and removed while the $H_2S$ is reacted with the remaining $SO_2$ in a Claus reactor to form additional sulfur. The tail gas from the Claus plant is incinerated and returned to the FGD system.

The Wellman-Lord system, because of the high solubility of the sodium liquor, is very effective in removing $SO_2$. All of the commercial utility systems are designed for 90 percent removal. The chemistry of the system is also very resistant to upsets. Sodium sulfite, as it reacts with $SO_2$ in the absorber, forms the more soluble sodium bisulfite. Consequently, scaling cannot occur as in calcium based systems. The high liquid phase alkalinity allows the system to operate at very low L/G ratios. Typically 3-4 trays are used in the absorber. Internal recirculation is used on the trays to maintain the proper liquid depth.

Commercial Designs—

Table 52 presents design parameters of the commercial Wellman-Lord systems. All of these units have been designed by Davy-McKee, the process vendor. The San Juan units have a spare absorber module in order to comply with the stringent New Mexico $SO_2$ standard. The Delaware City units produce steam and electricity for Getty Refining and Marketing company and burn petroleum coke, with an average sulfur content of 7.0 percent and fuel oil or refinery gas (81). The Mitchell unit was the first coal-fired Wellman-Lord installation and has been extensively tested under EPA sponsorship (82,83,84).

TABLE 52. WELLMAN-LORD FGD SYSTEM DESIGN

| Plant | Startup | Regulatory Status[a] | Percent Sulfur | L/G (gal/1000acf) | Trays | Unit Removal (%) | By-product |
|---|---|---|---|---|---|---|---|
| San Juan 1 | 4-78 | C | 0.8 | 1.7 | 5 | 90 | Sulfur |
| San Juan 2 | 8-78 | C | 0.8 | 1.7 | 5 | 90 | Sulfur |
| San Juan 3 | 12-79 | C | 0.8 | 1 | 5 | 90 | Acid |
| San Juan 4 | 6-82 | C | 0.8 | - | - | 90 | Acid |
| Dean H. Mitchell 11 | 7-76 | E | 3.5 | - | 3 | 90 | Sulfur |
| Delaware City 1 | 5-80 | E | 7.0 | 1.8 | 4 | 90 | Acid |
| Delaware City 2 | 5-80 | E | 7.0 | 1.8 | 4 | 90 | Acid |
| Delaware City 3 | 5-80 | E | 7.0 | 1.8 | 4 | 90 | Acid |

[a] Regulatory Class: C-Standard more stringent than 1979 NSPS; E-Standard equal al to or less than 1971 NSPS.

Source: Reference 2.

All of the units have been designed for 90 percent $SO_2$ removal. The number of trays required to obtain the desired removal is inversely proportional to the $SO_2$ concentration. At low sulfur levels, the driving force (ratio of the gas phase $SO_2$ concentration to the liquid phase equilibrium partial pressure of $SO_2$ above the absorbing liquor) is similar for both low and high sulfur applications because the $SO_2$ recovery system requires a highly concentrated solution of sodium bisulfite. At higher driving forces, more removal occurs at each stage, consequently, the desired percentage reduction is obtained in fewer stages.

The overall L/G ratio is very low due to the high sorption capacity of the liquor. Each tray uses internal recirculation in order to maintain an adequate level of liquid for contacting the flue gas.

As mentioned earlier, Wellman-Lord systems have been selected at the utility installations for site specific reasons. The San Juan units, are located on an Indian reservation in New Mexico. No land was available for throwaway sludge disposal. The Delaware City units may also have land availability problems. The Getty refinery handles the conversion to sulfuric acid, which is then used inside the plant and also sold commercially. The Mitchell unit was selected to demonstrate the technology, with cost sharing from the EPA.

The choice of the end by-product depends on local market conditions. Since sulfuric acid has a low bulk value, it cannot be economically shipped long distances. Sulfur is easier to store and is more versatile in its applications. However, the cost of natural gas to reduce $SO_2$ was a significant factor in the decision to produce acid rather than sulfur at San Juan 3 and 4 (82). Natural gas prices are also spurring interest in coal based reduction technology by Allied and Foster Wheeler (16).

Other areas of process improvements are also being examined by the vendor for utility and industrial applications. These areas include: minimizing the evaporate energy requirement and absorber pressure drop, optimizing the high-temperature purge treatment, investigation of oxidation inhibitors and converting sodium sulfate into caustic or soda ash.

Economics—

Table 53 presents several estimates of the cost of new Wellman-Lord systems. Estimates 1-3 were done by TVA using their old economic and design basis. In comparison, the similar performing limestone system cost $98 per KW, approximately 30-40 percent less than the Wellman-Lord systems. The acid producing plant has both lower capital and operating costs than the sulfur units. It should be noted that both of the sulfur units are costed using coal-reducing processes which have not been commercially demonstrated in the U.S.

Estimate 4 by Beychok is close to the capital cost of the previous estimates. The spare absorber offsets the effects of 2 years of inflation. The annual costs are different due to a different plant operating profile. The old TVA basis used a graduated profile which declined over the years whereas EPRI uses a constant load factor. The constant factor tends to lower the first year costs while raising the levelized charges. Estimates 5 and 6 by Bechtel are for high and low sulfur coal applications respectively. The low sulfur case had reduced capital costs because of smaller sulfur recovery facilities. The utilities and raw material requirements are lower for the same reason.

The total capital cost of Estimates 7 and 8 by Stearns-Roger appear to be inconsistent with the other costs shown. The installed capital costs for flue gas scrubbing, reheat handling, purge treatment and $SO_2$ regeneration in estimate 7 are 75 percent higher than the same charges in estimate 5. However, both of these studies used identical time frames and similar design basis. Estimates 5 and 6 were obtained by modifying a Davy-McKee bid for a 500 MW unit. Estimates 7 and 8 were made using detailed equipment costs and a standardized process configuration which was significantly different from the usual Wellman-Lord design.

TABLE 53. ESTIMATES OF NEW WELLMAN-LORD FGD SYSTEMS

| Estimate Number | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 |
|---|---|---|---|---|---|---|---|---|
| Source | TVA | TVA | TVA | Beychok | Bechtel | Bechtel | Stearns-Roger | Stearns-Roger |
| Sponsor | EPA | EPA | EPA | EPRI | EPRI | EPRI | EPRI | EPRI |
| Reference | 16 | 16 | 16 | 17 | 10 | 10 | 12 | 12 |
| Plant Size (MW) | 500 | 500 | 500 | 500 | 2 x 500 | 2 x 500 | 500 | 500 |
| Coal Type | Eastern | Eastern | Eastern | Eastern | Eastern | Western | Eastern | Eastern |
| Coal Sulfur (%) | 3.5 | 3.5 | 3.5 | 3.5 | 4.0 | 0.48 | 4.0 | 4.0 |
| NSPS | 1971 | 1971 | 1971 | 1971 | 1978 | 1978 | 1979 | 1979 |
| Base Year - Capital | 1979 | 1979 | 1979 | 1977 | 1978 | 1978 | 1978 | 1978 |
| Base Year - Operations | 1980 | 1980 | 1980 | 1977 | 1978 | 1978 | 1978 | 1978 |
| Absorber Removal (%) | 79 | 79 | 79 | 90 | 84 | 83 | 90 | 90 |
| Spare Module | No | No | No | Yes | Yes | Yes | Yes | Yes |
| Byproduct Type | Acid | Sulfur[a] | Sulfur[b] | Sulfur | Sulfur | Sulfur | Sulfur | Acid |
| Byproduct Credit | Yes | Yes | Yes | No | No | No | Yes | Yes |
| Capital Cost ($/KW) | 131 | 138 | 141 | 131 | 138 | 141 | 248 | 267 |
| Revenue Requirement (mills/KWh) | | | | | | | | |
| First Year | 5.11 | 6.03 | 5.94 | 7.32 | 10.4 | 6.5 | 11.9 | 10.5 |
| Levelized | 6.60 | 7.59 | 7.54 | 10.53 | 15.0 | 9.1 | 19.0 | 15.7 |

[a] Using the Resox Process

[b] Using the Allied coal process

[c] Using the Allied gas process

[d] 1978 Proposed NSPS

Costs reported to and adjusted by PEDCo for commercial systems are presented in Table 54. The adjusted capital costs are significantly higher than any of the previously described processes. The Mitchell unit is fairly small which normally results in a high dollar per KW charge in most cost studies. The San Juan units have a significant amount of spare capacity in the absorber module, main evaporators, purge treatment and sulfur units. This was done in order to achieve essentially 100 percent reliability (82).

## Magnesium Oxide

Magnesium oxide, the last FGD system defined as commercially used, is currently being constructed at 3 units owned by the Philadelphia Electric Company (PECo). These units are all being retrofit based on a consent decree which is not as stringent as the 1979 NSPS. The FGD systems are scheduled to start up in 1983. The magnesium oxide process has been tested at three utility sites. All of these units have been dismantled.

### Process Description--

The magnesium oxide process flow diagram is shown in Figure 11. Flue gas is scrubbed in a precooler to remove chlorides. It is then contacted with a slurry of magnesium sulfite in the absorber. Some oxidation of sulfite to sulfate occurs, along with precipitation of the two salts. The following reactions occur.

$$\begin{aligned}
SO_2(g) + H_2O(1) + MgSO_3(1) &\rightleftarrows Mg(HSO_3)_2 \\
MgSO_3(1) + 1/2\ O_2(g) &\rightleftarrows MgSO_4(1) \\
MgSO_3(1) + 6H_2O &\rightleftarrows MgSO_3 \cdot 6H_2O(s) \\
MgSO_3(1) + 3H_2O &\rightleftarrows MgSO_3 \cdot 3H_2O(s) \\
MgSO_4(1) + 7H_2O &\rightleftarrows MgSO_4 \cdot 7H_2O\ (s)
\end{aligned}$$

## TABLE 54. OPERATIONAL WELLMAN-LORD SYSTEM COSTS

| Unit | Capital Costs ($/KW) |  | Annual Costs (mills/KWh) |  |
|---|---|---|---|---|
|  | Reported | Adjusted | Reported | Adjusted |
| D. H. Mitchell 11 | 157.4 | 254.6 | 13.02 | 20.79 |
| San Juan 1 | 132.8 | 277.9 | - | 16.69 |
| San Juan 2 | 137.1 | 282.2 | - | 16.89 |
| Average | 142.4 | 271.6 | - | 18.12 |
| Std. Dev. | 13.1 | 14.9 | - | 2.31 |

Source: Reference 2

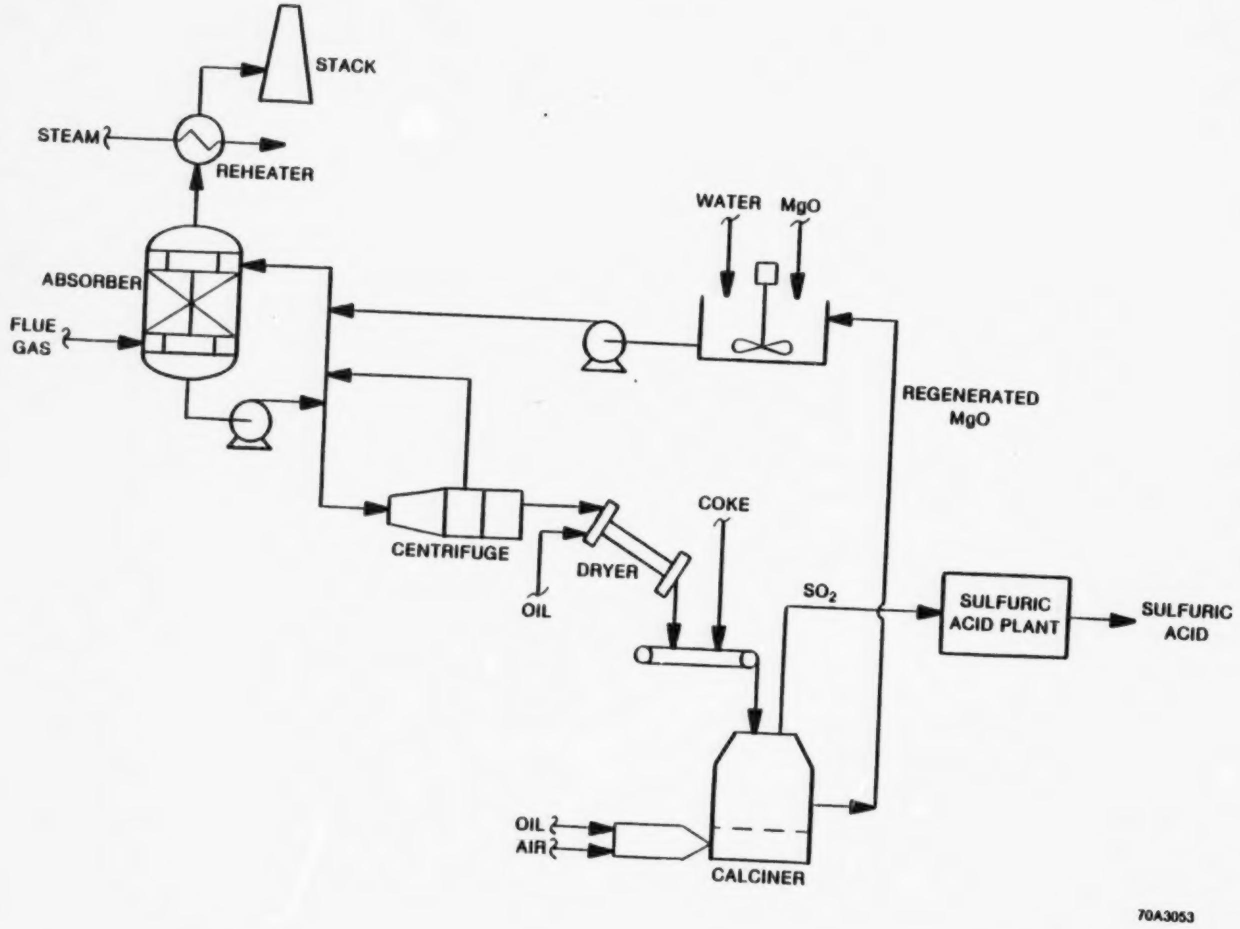

STACK
STEAM
REHEATER
ABSORBER
FLUE GAS
WATER
MgO
REGENERATED MgO
CENTRIFUGE
DRYER
OIL
COKE
$SO_2$
SULFURIC ACID PLANT
SULFURIC ACID
OIL
AIR
CALCINER
70A3053

Figure 11. Magnesium Oxide Process Flow Diagram

A portion of the slurry is withdrawn and centrifuged. The solids are dried further while the liquid is returned to the absorber. In the dryer, the waters of hydration of both magnesium sulfite and sulfate are removed. Regeneration of the solids is accomplished in a calciner, where the following reactions occur. Coke is added to reduce the magnesium sulfate present.

$$MgSO_3(s) \rightarrow MgO(s) + SO_2(g)$$
$$MgSO_4(s) + 1/2C(s) \rightarrow MgO(s) + SO_2(g) + 1/2CO_2(g)$$

The $SO_2$ is present in the calciner off-gas at 7-10 percent which is dried and sent to a sulfuric acid plant.

The regenerated magnesium oxide is returned to the absorber loop. Makeup MgO is added as needed and the mixture is slaked to produce magnesium hydroxide.

Several problem areas have been identified in the process through the operation of prototype units. As shown in the first set of reactions, both the tri-and hexahydrate crystal can form. The trihydrate crystal is thermodynamically preferred at normal operation temperatures (85). During startup and shutdown though the hexahydrate can form. The two types of crystals cause handling problems because of the differences in their size and shape. Equipment designed for one type of crystal often experienced operating problems when the other crystal shape was presented in the pilot tests.

Problems have also been experienced in the calciner, where magnesium oxide is regenerated. The reduction of magnesium sulfate with coke occurs at about 1600-1800°F. Several units have experienced dead burning of the solids, creating large deposits in the furnace. Chlorides present in the system are also converted to HCl in the calciner and can cause corrosion problems.

Recent Innovations--

Three major demonstrations of magnesium oxide scrubbing have been conducted in the past 10 years. These test programs have demonstrated the technical viability of the technology and have resulted in many process design improvements. The first unit was at the Mystic Station of Boston Edison, operated from 1971-1974 (86). This 150 MW station fired 2.5 percent sulfur oil. The FGD system was designed by Chemico, now a part of General Electric Environmental Services Inc. The first coal fired test plant was at the Dickerson station of Potomac Electric and Power (87). This 95 MW station was also built by Chemico and operated from 1973-1975.

The longest demonstration program has been conducted by United Engineers and Philadephia Electric at the Eddystone station from 1975-1978. This demonstration program was performed in order to obtain design information on the system (88,89). Full-scale units are in construction for both Eddystone units and one unit at the Cromby station of PECo.

Several process improvements have been investigated with a 1 MW pilot unit and at the 10 MW Shawnee prototype by TVA (90,91). These improvements include using a spray dryer and fabric filter instead of a slurry, using coal or a coal-oil mixture for the drying and calcining of the magnesium sulfite, and the direct production of elemental sulfur by decomposing $MgSO_3$. The tests at Shawnee investigated the use of the cocurrent absorber design. At one time, TVA was considering construction of a 600 MW unit at the Johnsonville steam plant (3). Those plans have been dropped and low sulfur coal is being used at this site (92).

Economics--

Although currently there are no magnesium oxide plants in commercial operation, process economics have been estimated by a number of investigators. Several of these estimates are presented in Table 55. All of the

### TABLE 55. ESTIMATES OF NEW MAGNESIUM OXIDE FGD SYSTEMS

| Estimate Number | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| Source | TVA | TVA | TVA | Bechtel | Bechtel | Beychok | SRI |
| Sponsor | EPA | EPA | EPA | EPRI | EPRI | EPRI | EPRI |
| Reference | 4 | 4 | 4 | 10 | 10 | 17 | 11 |
| Plant Size (MW) | 500 | 500 | 500 | 2 x 500 | 2 x 500 | 500 | 500 |
| Coal Type | Eastern | Eastern | Eastern | Eastern | Western | Eastern | Eastern |
| Coal Sulfur (%) | 3.5 | 2.0 | 5.0 | 4.0 | 0.48 | 3.5 | 3.5 |
| NSPS | 1971 | 1971 | 1971 | 1978[a] | 1978[a] | 1971 | 1971 |
| Base Year-Capital | 1979 | 1979 | 1979 | 1978 | 1978 | 1977 | 1979 |
| Base Year-Operations | 1980 | 1980 | 1980 | 1978 | 1978 | 1977 | 1979 |
| Absorber Type | Grid | Grid | Grid | Venturi | Venturi | Venturi | TCA |
| Absorber Removal (%) | 79 | 63 | 85 | 85 | 83 | 90 | 93 |
| Spare Module | No | No | No | Yes | Yes | Yes | Yes |
| L/G (gallons/1000 acf) | 10 | 10 | 10 | 50 | 25 | - | 16 |
| Capital Cost ($/KW) | 132 | 108 | 152 | 176 | 129 | 136 | 193 |
| Revenue Requirements (mills/KWh) | | | | | | | |
| First Year | 5.08 | 4.19 | 5.83 | 9.2 | 6.3 | 7.4[b] | 6.9[b] |
| Levelized | 6.60 | 5.42 | 7.57 | 12.7 | 8.6 | 10.3[b] | 12.1[b] |

[a] 1978 Proposed NSPS

[b] Does not include sulfuric acid credit

estimates include a sulfuric acid plant to process the $SO_2$ stream. Estimates 1-3 are by TVA using their old design and economic basis. In comparison to estimate 1, the same limestone system in Table A7 costs $98 per KW with a levelized revenue requirement of 5.14 mills per KWh. As is expected, higher sulfur levels increase both capital and operating costs.

Estimates 4 and 5 by Bechtel are for a high and low sulfur coal respectively. These costs are significantly higher because of the spare module and 100 percent bypass ducts around the absorbers. Estimate 6 is from a study of regenerable processes sponsored by EPRI. The capital cost in estimate 6 is for a system capable of meeting the 1977 NSPS, expressed in 1977 dollars. A similar limestone system in this study cost $94 per KW. The revenue requirements in this case did not include a credit for the sulfuric acid produced. In the other studies, this credit is on the order of 0.5-1.0 mill per KWh. Estimate 7 is from a third EPRI report examining the costs of high removal efficiency systems. This estimate includes a spare module and bypass ductwork and has the highest capital cost of all the cases. Credit for the sulfuric acid byproduct was not included in the annual costs.

## DEMONSTRATION FGD PROCESS DESCRIPTIONS

This section discusses several of the emerging FGD technologies which have been tested recently or are undergoing further development. A brief description of each process along with their current developmental status is provided. Economics are not presented due to the lack of commercial scale operating experience of most of these processes. Processes considered include: Aqueous Carbonate, Citrate, Chiyoda Thoroughbred 121 (CT-121), DOWA, dry injection, electron-beam, and Shell copper oxide.

### Aqueous Carbonate

The Aqueous Carbonate process is a regenerable spray drying system using sodium carbonate as the scrubbing reagent. Developed by Rockwell International, the Aqueous Carbonate system is currently in the prototype stage. All of the process steps have been individually tested on a 1000 scfm scale or are considered by the developer to be proven technology. The spray dryer has been tested in two pilot programs; one at Bowen Engineering's North Branch, New Jersey test facility and the other at Southern California Edisons' Mohave Station (93). The regeneration section has recently been tested on a 100 MW unit at Niagra Mohawk Power Corporation's Huntley Electric Generating Station using a synthetic reaction product. The objectives of these tests were to use coal instead of petrolum coke for regeneration, simulate integrated operation by operating the regeneration steps sequentialty, and support the commercial size demonstration of the process by simulating reduction and carbonation steps and evaluating construction materials (94). These objectives were accomplished although the addition of coal did make the process more difficult to operate and control. An integrated 100 MW demonstration unit at the Huntley Station is scheduled for startup in 1982. Testing at Huntley is cosponsored by Niagara Mohawk Power Corporation and the Electric Power Research Institute (94).

A flow schematic of the Aqueous Carbonate process is shown in Figure 12. After fly ash removal, the flue gas enters a spray dryer scrubber at temperatures between 250 and 350°F. In the scrubber, flue gas contacts atomized droplets of $Na_2CO_3$ solution. These droplets are generated by high speed centrifugal atomizers. $SO_2$ is sorbed into the liquor and converted to sodium sulfite by the reaction:

$$SO_2(g) + Na_2CO_3(1) \rightarrow Na_2SO_3(s) + CO_2(g).$$

Typically, 25 percent of the sulfite is oxidized to sulfate by the reaction:

$$Na_2SO_3(1) + 1/2O_2(g) \rightarrow Na_2SO_4(s).$$

Also, some $SO_3$ is removed in the absorber:

$$SO_3(g) + Na_2CO_3(1) \rightarrow Na_2SO_4(s) + CO_2(g).$$

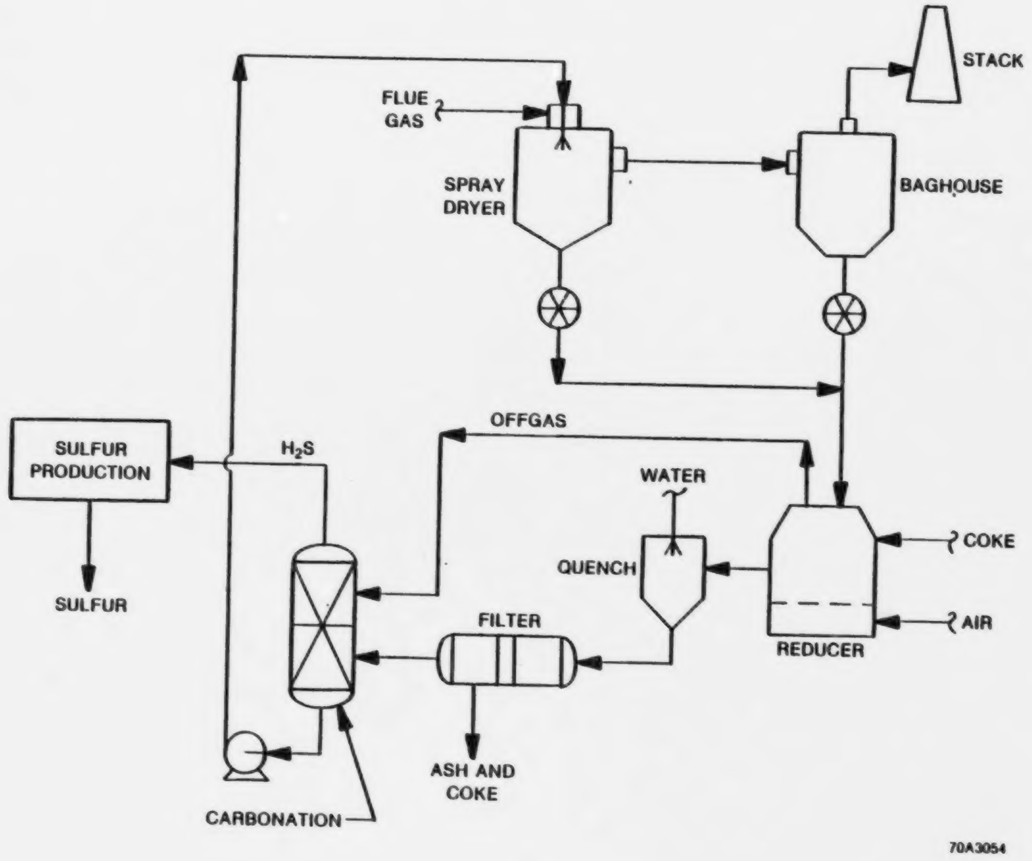

STACK
FLUE GAS
SPRAY DRYER
BAGHOUSE
OFFGAS
$H_2S$
SULFUR PRODUCTION
SULFUR
WATER
QUENCH
COKE
AIR
REDUCER
FILTER
ASH AND COKE
CARBONATION
70A3054

Figure 12. Aqueous Carbonate Process Flow Diagram

Liquid-to-gas ratios are about 0.3 gallons per thousand standard cubic feet of gas. Because of this low liquid rate, insufficient water is present to saturate the gas, and the spent reactant is entrained as dry particles. The flue gas is typically cooled to 150-175°F. The dust-laden flue gas is routed to a bank of product collection cyclones where the majority of the dry particles are removed. Additional particulate removal is accomplished in high efficiency electrostatic precipitators. The dry product collected in the cylcones and precipitators is stored in a silo and then conveyed pneumatically to the reducer vessel. This vessel contains a pool of molten salts at a high temperature (1700-1900°F). Carbon is injected as petroleum coke or coal which regenerates the spent reactant by the following reactions:

$$Na_2SO_3(l) + 3/2C(s) \rightarrow Na_2S(l) + 3/2CO_2(g)$$
$$Na_2SO_4(l) + 2C(s) \rightarrow Na_2S(l) + 2CO_2(g)$$
$$C(s) + O_2(g) \rightleftarrows CO_2(g)$$

The last reaction is exothermic and provides heat to maintain the system's temperature. The $CO_2$ rich off-gas is sent to the carbonation section. An aqueous purge stream is also discharged with the cooling water in the reduction section to prevent exessive buildup of chlorides. The sodium sulfide melt is dispersed into fine droplets by the reducer discharge nozzles (steam shatter jet) and dissolved in a solution (called green liquor) near its boiling point. Insoluble fly ash and unreacted coke or coal are removed in a rotary drum filter. After filtration, the green liquor is contacted with the $CO_2$ rich reducer off-gas, producing a concentrated $Na_2CO_3$ solution for recycle to the scrubber and an $H_2S$ rich off-gas for feed to the sulfur plant according to the reactions:

$$Na_2S(l) + H_2O + CO_2(l) \rightarrow Na_2CO_3(l) + H_2S(g)$$
$$Na_2CO_3(l) + H_2O + CO_2 \rightleftarrows 2NaHCO_3(s)$$

The process currently uses a Claus plant for conversion of $H_2S$ to elemental sulfur (94).

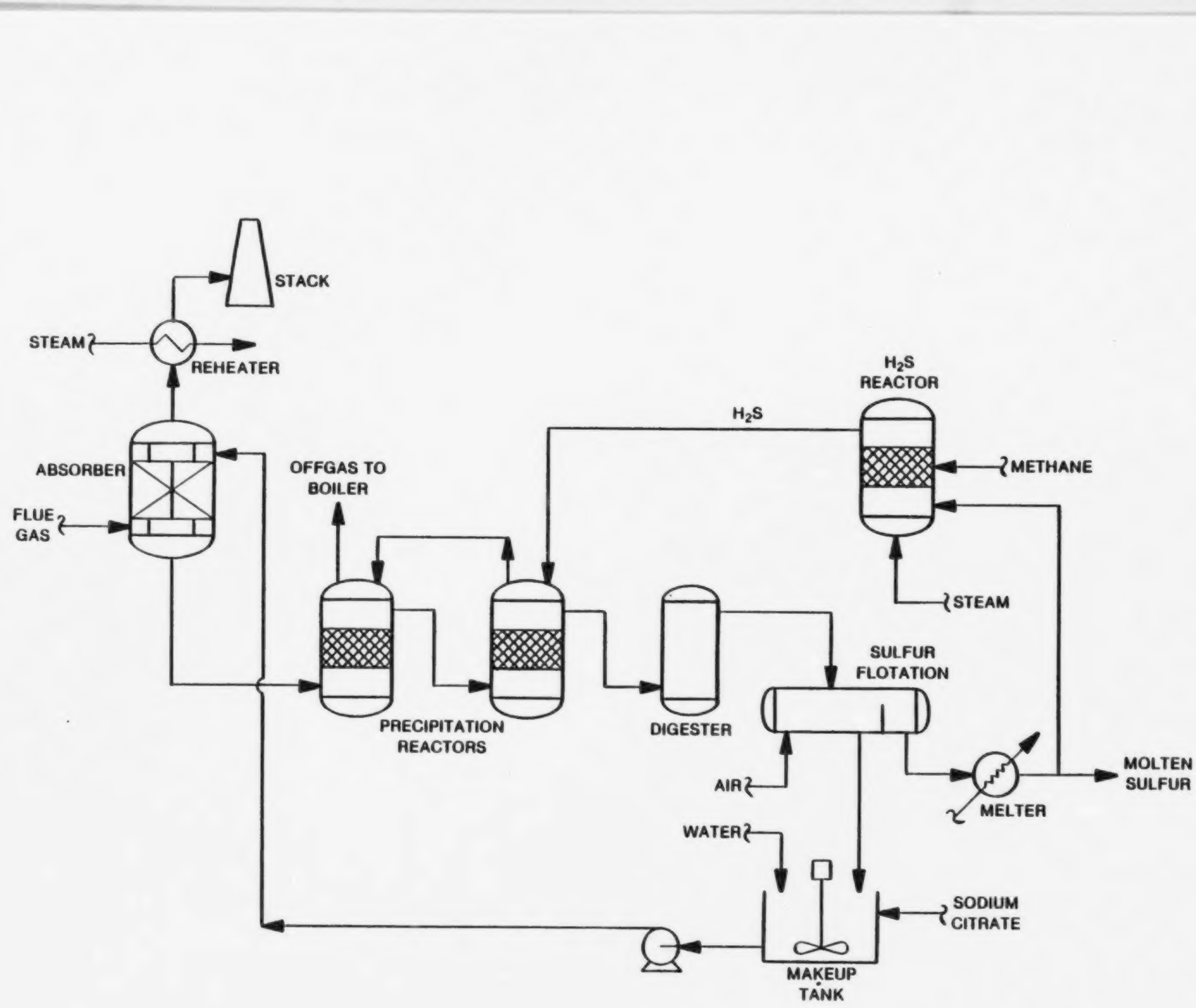
## Citrate

The Citrate process uses a buffered aqueous solution of sodium citrate to remove $SO_2$ from flue gases. Absorbed $SO_2$ is reduced to elemental sulfur by reaction with gaseous $H_2S$ or by steam stripping with subsequent processing of the $SO_2$-rich vapors. Three U. S. pilot plant operations have tested the $SO_2$ removal ability of the process. The U. S. Bureau of Mines in conjunction with private industry has tested two pilot units, one at Magma Copper Company's copper smelter at San Manuel, Arizona and another at Bunker Hill company's lead smelter at Kellogg, Idaho. A third pilot unit was also tested by Arthur G. McKee and Company, Peabody Engineering Systems, and Pfizer Inc. on a 2000 scfm stream from a coal-fired steam generating station in Terre Haute, Indiana. These pilot plant tests have shown the Citrate process capable of $SO_2$ removals of 95 percent and better. Operating problems were encountered, however, generally associated with equipment pluggage due to sulfur condensation (95,96,97,98).

The Citrate process is being commercially applied to a refinery tail gas (99). In addition, a 60-MW coal-fired demonstration plant, sponsored by the U. S. Bureau fo Mine, has been constructed at St. Joe Mineral Corporation's George F. Weaton power plant at Monaca, PA. Integrated operation of the demonstration unit, however, has not been accomplished to date (100, 101).

A simplified flow schematic of the most developed version of the process is shown in Figure 13. After passing through a high efficiency particulate removal device, the flue gas is quenched in a prescrubber to remove fly ash, $SO_3$, and chlorides and saturate the gas. The flue gas is then sent through an absorber where $SO_2$ is sorbed with sodium citrate according to the reactions:

$$SO_2(g) + H_2O \rightleftarrows H_2SO_3^- + H^+$$
$$HCit^= + H^+ \rightleftarrows H_2Cit^-$$


STACK
STEAM
REHEATER
$H_2S$ REACTOR
$H_2S$
ABSORBER
OFFGAS TO BOILER
METHANE
FLUE GAS
STEAM
SULFUR FLOTATION
PRECIPITATION REACTORS
DIGESTER
AIR
MOLTEN SULFUR
MELTER
WATER
SODIUM CITRATE
MAKEUP TANK

70A3055

Figure 13. Citrate Process Flow Diagram

As shown here, the citrate ion acts as a pH buffer to maintain the scrubber pH at 3.5 to 5.0 The ability to neutralize hydrogen ions in the scrubbing liquor aids the sorption reaction.

After absorption, the $SO_2$-rich sorbent is routed to the regeneration unit. A number of alternatives have been proposed for the regeneration system. The most developed scheme is shown in Figure 11. Here, an $H_2S$-rich gas is used to produce elemental sulfur in the regeneration reactors. The regeneration reaction sequences is complicated but is basically a liquid phase Claus reaction. Overall, the regeneration reaction can be described as:

$$SO_2(1) + 2H_2S(g) \rightleftarrows 3S(s) + 2H_2O$$

Recent investigation by the Bureau of Mines indicate that the rate-limiting reaction step is (102):

$$2H_2S(g) + S_2O_3^{=} + 2H^{+} \rightleftarrows 4S(s) + 3H_2O$$

Sulfur formed in the regenerator is separated from the lean sorbent liquor in a sulfur flotation unit. The lean sorbent liquor is fortified by the addition of citric acid or sodium citrate, pH adjusted with NaOH or sodium carbonate, and returned to the absorber. The sulfur slurry is sent to a sulfur melter where molten sulfur is produced at 280°F. The sulfur is then separated from the entrained liquor in a decanter. The decanted liquor is returned with the absorber blowdown to the regeneration unit. The sulfur product is stored for disposal or sale. If there is not an available source of $H_2S$, some of this sulfur may be used for $H_2S$ production.

Both Flakt-Boliden and Peabody Engineering have also proposed steam stripping for regeneration (16,103,104). A flow diagram of the Peabody process combined with a RESOX system is shown in Figure 14. In this version, the $SO_2$ rich sorbent is stripped with live steam in a counter current column. The $SO_2$ rich overhead is typically 95 percent $SO_2$ and is suitable for

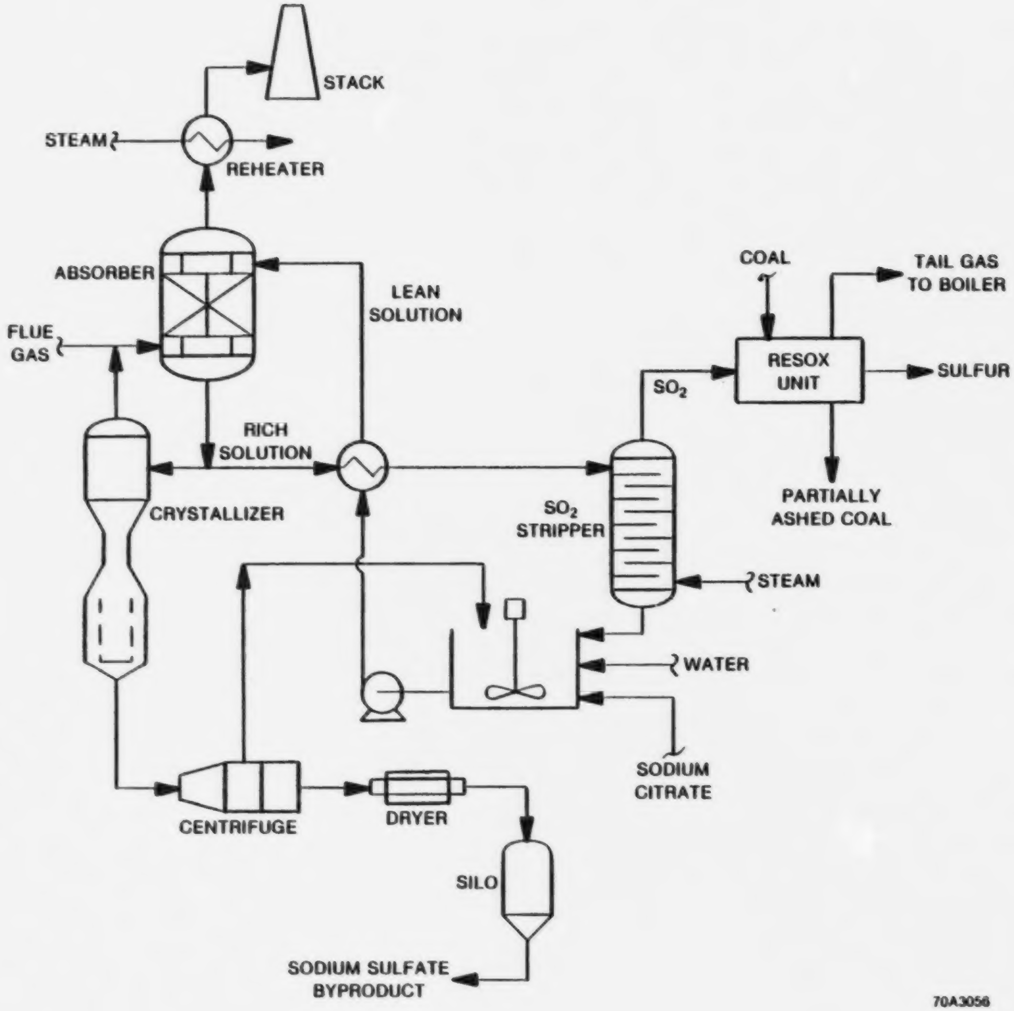

STACK
STEAM
REHEATER
ABSORBER
FLUE GAS
LEAN SOLUTION
RICH SOLUTION
CRYSTALLIZER
COAL
TAIL GAS TO BOILER
RESOX UNIT
SULFUR
$SO_2$
PARTIALLY ASHED COAL
$SO_2$ STRIPPER
STEAM
WATER
SODIUM CITRATE
CENTRIFUGE
DRYER
SILO
SODIUM SULFATE BYPRODUCT
70A3056

Figure 14. Citrate-Resox Process Flow Diagram

conversion to sulfuric acid, elemental sulfur, or liquified $SO_2$. This process has been tested at the pilot plant level.

One of the major operational problems with the Citrate process is the presence of sulfates in the scrubbing liquor. Sulfur trioxide is also directly absorbed from the flue gas. Sulfate can also be formed in the scrubbing liquor by the following reactions:

$$HSO_3^- + 1/2O_2(g) \rightarrow HSO_4^-$$
$$3S_2O_3^= + 2H^+ \rightarrow 2SO_4^= + H_2O + 4S(s)$$

Typically, sulfite oxidation is on the order of one to four percent of the $SO_2$ removed. In most of the Citrate processes, the sulfate is removed as $Na_2SO_4 \cdot 10H_2O$ by crystallization (67). Other problems identified to date involve integration of the $H_2S$ production process, pluggage due to sulfur condensation, and development of the stripping alternative.

## Chiyoda Thoroughbred-121

The Chiyoda Thoroughbred-121 (CT-121) process is a wet limestone scrubbing process developed by Chiyoda Chemical Engineering and Construction Company. The CT-121 system combines the reactions of the earlier Chiyoda 101 process in a single vessel while using a novel gas-liquid contacting mechanism. Following experimental tests on Japan, a 23 MW prototype system was successfully demonstrated at Gulf Power Company's Scholz Generating Station (105). This ten month evaluation program, sponsored through an agreement by Chiyoda, EPRI, and Southern Company Service, demonstrated to viability of the process on a coal-fired boiler. Chiyoda has sold a CT-121 system for an 80 MW equivalent oil-fired industrial boiler in Japan and is actively marketing this process in the U. S. for both utility and industrial applications.

The key to this process is the Jet Bubbling Reactor (JBR) in which $SO_2$ removal, oxidation, and precipitation occur simultaneously (106). This vessel is shown in Figure 15. The overall reaction can be described as:

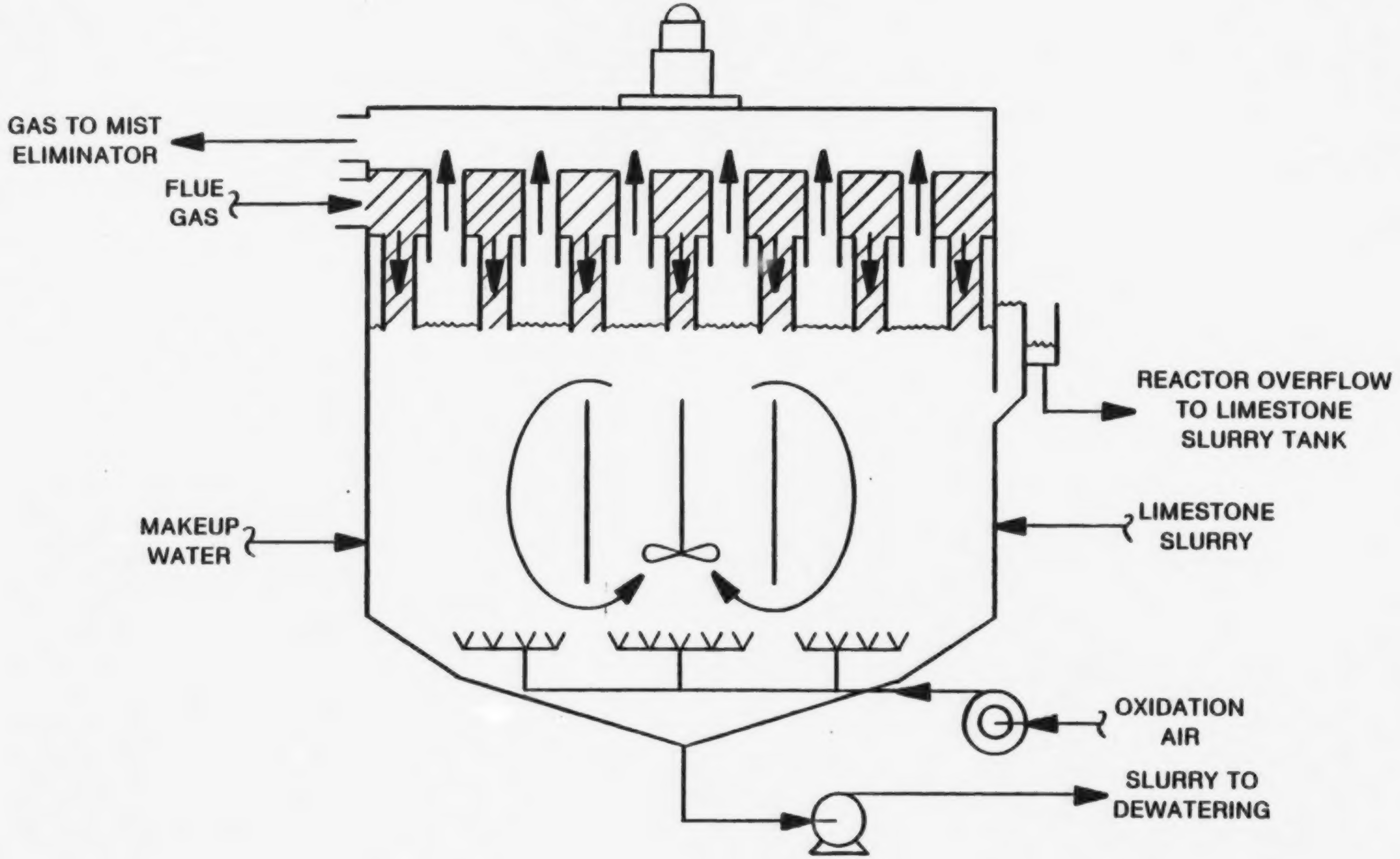

GAS TO MIST ELIMINATOR
FLUE GAS
MAKEUP WATER
REACTOR OVERFLOW TO LIMESTONE SLURRY TANK
LIMESTONE SLURRY
OXIDATION AIR
SLURRY TO DEWATERING
70A3057

Figure 15. Schematic of Jet Bubbling Reactor

$$SO_2(g) + CaCO_3(s) + 1/2O_2(g) + 2H_2O \rightarrow CaSO_4\cdot 2H_2O(s) + CO_2(g)$$

Flue gas enters the JBR through spargers located below the slurry surface. The sparger immersion depth can be controlled with a adjustable weir, but usually ranges from 9 to 13 inches for 90 percent $SO_2$ removal (107).

Air is introduced at the bottom of the JBR to oxidize the absorbed sulfite to sulfate forming calcium sulfate (gypsum) crystals. Low pH operation, typically 3.5 to 4.5, helps promote the oxidation reaction. However, at these low pH values it is very important that complete oxidation be achieved since any sulfite not oxidized will quickly raise the $SO_2$ back pressure and limit the $SO_2$ removal.

Another advantage of the low pH operation is improved limestone utilization. An average limestone utilization of 97 percent was reported during the prototype demonstration program at Scholz (107). Slurry reentrainment should be low with this process reducing the mist eliminator loading. Slurry which is reentrained should also be low in $CaCO_3$ due to the high limestone utilizations. Excess $CaCO_3$ can be a contributor to mist eliminator fouling. No problems were experienced with the mist eliminators during the demonstration program.

### Dowa

The Dowa process uses basic aluminum sulfate to remove $SO_2$. Dowa Mining Company began developing this process in the early 1970s with the first commercial unit completed in October 1972. A total of seven plants in Japan and one in China currently use the DOWA process (42).

The first coal-fired application of the process has been tested at the Shawnee test facilities at TVA. The results from these tests show that the DOWA system is capable of greater than 90 percent $SO_2$ removal for a coal-generated flue gas. Some differences in operation, notably in the oxidation rate were observed at Shawnee compared to the Japanese installations. The

Shawnee work was sponsored by EPRI, TVA, and UOP with funding support from the Environmental Protection Agency (108,109). UOP is commercially marketing this process (110).

The Dowa process uses a basic aluminum sulfate solution to absorb $SO_2$. This solution is oxidized and then regenerated by limestone addition, which forms gypsum. A process flow diagram is shown in Figure 14. In the absorber, the $SO_2$ is removed by the solution as shown below.

$$Al_2(SO_4)3 \cdot Al_2O_3(1) + 3SO_2(1) \rightleftarrows Al_2(SO_4)3 \cdot Al_2(SO_3)_3(1)$$

This liquor is sent to the oxidizer vessel where air oxidizes the sulfite to sulfate.

$$Al_2(SO_4) \cdot Al_2(SO_3)_3(1) + 3/2O_2(1) \rightarrow 2Al_2(SO_4)_3(1)$$

A portion of this liquor is neutralized by limestone while the majority is recycled back to the absorber. The neutralization reaction regenerates the basic aluminum sulfate and precipitates gypsum.

$$\begin{aligned} 2Al_2(SO_4)_3(1) + 3CaCO_3(1) + 6H_2O \rightleftarrows\ & Al_2(SO_4)_3 \cdot Al_2O_3(1) + \\ & 3CaSO_4 \cdot 2H_2O(s) + 3CO_2(g) \end{aligned}$$

The gypsum is separated from the liquor by thickening and centrifuging. Makeup aluminum sulfate is added as needed to maintain an aluminum concentration of 20 grams per liter. In the commercial Japanese plants, this corresponds to about one pound per ton of gypsum produced (111).

The Dowa process operates at a pH of approximately 3. Low pH operation may require more acid resistant materials of construction over conventional limestone scrubbing systems (109). However, chemical scaling should be less of a problem with the DOWA system as well as equipment erosion since it utilizes a clear scrubbing solution. During the tests at Shawnee calcuim sulfate scale deposits for previous limestone testing were dissolved during the

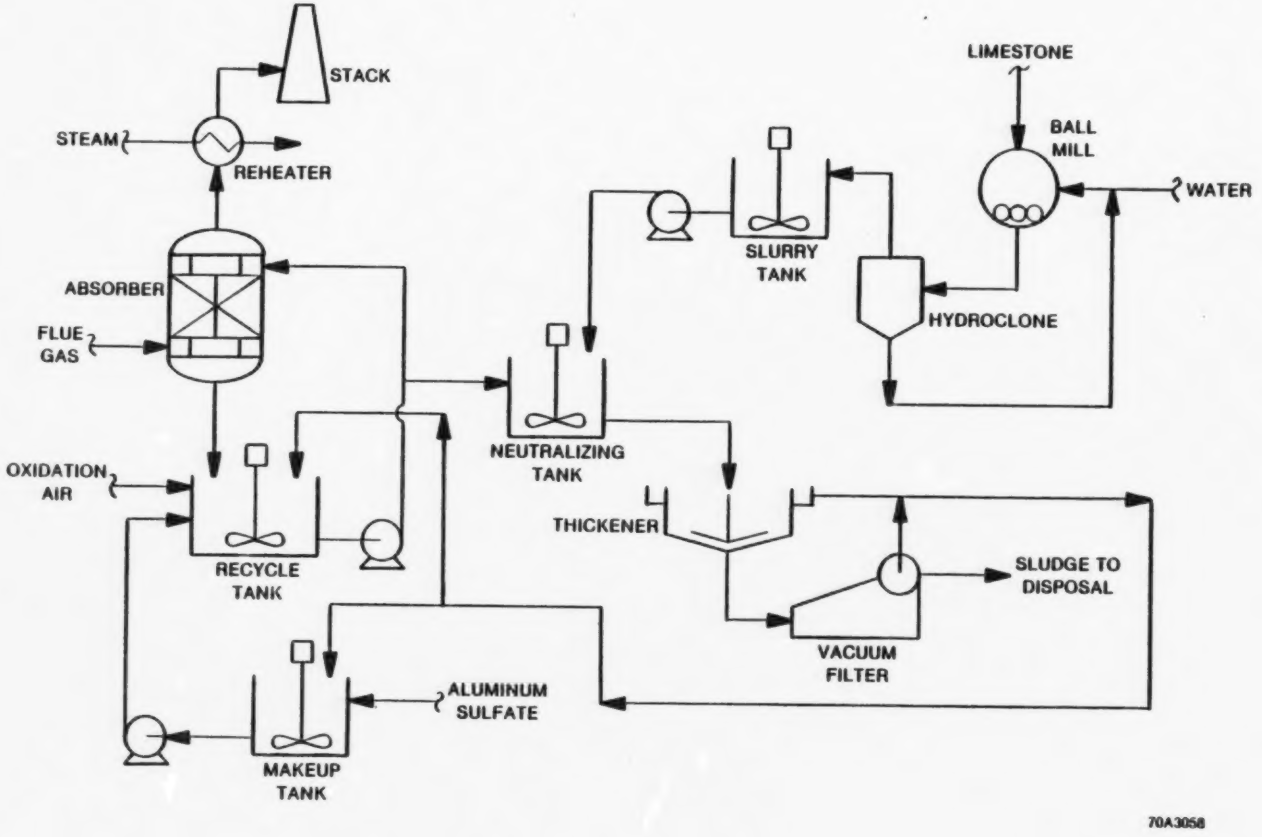

STACK
STEAM
REHEATER
ABSORBER
FLUE GAS
OXIDATION AIR
RECYCLE TANK
MAKEUP TANK
ALUMINUM SULFATE
NEUTRALIZING TANK
SLURRY TANK
LIMESTONE
BALL MILL
WATER
HYDROCLONE
THICKENER
SLUDGE TO DISPOSAL
VACUUM FILTER
70A3058

Figure 16. Dowa Aluminum Sulfate Process Flow Diagram

Dowa testing (110). The changeover to the Dowa system at Shawnee also required only minor process and equipment modifications. Therefore, the Dowa system could conceivably be attractive for upgrading existing limestone systems.

## Dry Injection

Dry injection of alkaline sorbents has been examined by EPA, DOE, and EPRI. A summary of past and current dry injection testing programs is shown in Table 56. The basis for the process is the injection of a dry powdered alkaline solid in the flue gas to remove $SO_2$. The solid particules are then removed in a downstream particulate removal device such as a baghouse. The reagent injection point can vary from the furance burners to before the particulate collector. One variation of the process is to precoat the baghouse bags with the sorbent.

At present, no commercial dry injection systems are in operation. However, several demonstration programs are currently underway or have been recently completed. The largest unit being tested is a 22 MW unit at Public Service of Colorado's Cameo Station. In all, as many as 11 utilities have involved in dry injection testing since 1965.

Perhaps the biggest problem with dry injection is reagent availability. Nahcolite, one of the two probable reagents, is currently not produced in commercial quantities. A development by Multi Minerals is scheduled to produce one million tons per year by 1986, which could supply reagent for about 5000 MW of low sulfur coal capacity. However, these nahcolite deposits are interspaced with oil shale, whose development may require a synfuels industry. Therefore, the availability of this nahcolite is not certain at the present time. Trona, the other probable reagent, may be more profitable to producers of soda ash, because of tax depletion allowances, rather than as an FGD reagent. The second problem limiting dry injection applications is operating costs. Although the capital cost of the injection-baghouse system is undoubtably less than a limestone system, the reagent cost differential is substantial. As the coal sulfur content increases, more reagent is

TABLE 56. DRY INJECTION $SO_2$ REMOVAL OPERATING EXPERIENCES

| System/Location | Developer and/or Supplier | Scale | Sorbent | Status | Comments |
|---|---|---|---|---|---|
| Southern California Edison Company | Babcock and Wilcox | Demonstration (320 MW) | Dolomitic limestone | Started: 1965<br>Ended : 1965 | Objective: $SO_2$ removal for plume opacity control |
| Alamitos Station (Los Angeles, CA) | | Test Unit (1500 ACFM w/2 bags) | Dolomitic limestone, nahcolite | Started: 1967<br>Ended : 1969 | Test results were encouraging, but the use of nahcolite was not considered further due to its unavailability as a firm source. |
| Public Service of Indiana - Edwardsport Station (Bicknell, IN) | Wheelabrator-Frye, Inc. | Pilot-Scale | $NaHCO_3$, soda ash, potassium permanganate, calcium hydroxide | Started: 1967<br>Ended : 1969 | $NaHCO_3$, and soda ash were effective |
| Public Service Electric & Gas of New Jersey - Mercer Station (Newark NJ) | Air Preheater Co., Inc. | Pilot-Scale | $NaHCO_3$, nahcolite, hydrated lime | Started: 1968<br>Ended : 1969 | Nahcolite and $NaHCO_3$ performed effectively |
| Colorado Ute Electric Assc. - Nucla Station (Nucla, CO) | Wheelabrator-Frye, Inc. | Full-Scale (11 MW) | Nahcolite | Started: 1974<br>Ended : 1974 | $SO_2$ removal efficiencies: 50-70% at inlet $SO_2$ concentrations of 400-500 ppm. |
| Otter Tail Power Company-Hoot Lake Station, Unit 3 (Fergus Falls, MN) | Wheelabrator-Frye, Inc. | Pilot-Scale | Nahcolite, CaO, $Ca(OH)_2$ | Started: 1974<br>Ended : 1974 | $SO_2$ removal efficiency: 46% at air-to-cloth ratio of 2.3/1 |
| Basin Electric Power Company-Leland Olds Station, Unit No.2 (Stanton, ND) | Wheelabrator-Frye, Inc. | Pilot-Scale | Nahcolite | Started: 1976<br>Ended : 1977 | $SO_2$ removal efficicies: 83-90%; sorbent utilization efficiencies: 60-77% |
| Colorado Springs Municipal Utility-Martin Drake Station, No. 6 Boiler (Colorado Springs, CO) | Envirotech Corp. Buell Emission Control Division | Pilot-Scale (4500 ACFM) | Nahcolite, raw trona, refined trona | Started: 1977<br>Ended : Ongoing | Results of dry injection tests showed nahcolite to be the best sorbent, followed by refined trona. Experiments are also being conducted to evaluate the disposal of the sodium sulfate/fly ash waste (SINTERNA process). |

(Continued)

TABLE 56. DRY INJECTION $SO_2$ REMOVAL OPERATING EXPERIENCES (Continued)

| System/Location | Developer and/or Supplier | Scale | Sorbent | Status | Comments |
|---|---|---|---|---|---|
| KVB Facility (Tustin, CA) | KVB/EPRI | Bench-Scale | $NaHCO_3$, nahcolite, trona | Started: 1977<br>Ended : 1978 | The tests have demonstrated the technical feasibility for a system firing a western coal |
| University of Tennessee (UT) Facility (Knoxville, TN) | UT/Carborundum | Pilot-Scale | $NH_3$, $NaHCO_3$ | Started: 1976<br>Ended : Ongoing | Both gaseous ammonia and $NaHCO_3$ were suitable additives with certain limitations |
| Kert Industries (Concord, NC) | EPA IERL-RTP | Pilot-Scale | -- | Started: 1979<br>Ended : Ongoing | Future study is being contemplated to include a regenerable sorbent process |
| Energy and Pollution Control, Inc., Facility (Bensonville, IL) | Energy and Pollution Control (EPC) of Flick-Reedy | Pilot-Scale | Lime, soda ash | Ongoing | $SO_2$ removal efficiencies ranging from 45 to 95% are accomplished by dry lime injection |
| DOE/Grand Forks Energy Technology | DOE/GFETC | Test Unit (200 ACFM) | Trona, nahcolite | Ongoing | Preliminary results indicate up to 90% utilization (based on $NO_x$ and $SO_2$) has been obtained with nahcolite and trona using western coal. Further tests on a new 130-scfm pulse jet baghouse are planned. |
| DOE/Pittsburgh Energy Technology Center Labs | DOE/GFETC | Test Unit (500 lb coal/hr. furnace) | Nahcolite, trona commerical $NaHCO_3$ | Started: 1979<br>Ended : 1980 | The tests showed nahcolite to be the best sorbent, obtaining up to 95% $SO_2$ removal with 1.1, 1.6, 3.5% S coals. |
| Public Service of Colorado - Cameo | EPRI/KVB/Mutli Mineral | Full-Scale (22 MWe) | Nahcolite | Started: 1980<br>Ended : Ongoing | |

Source: References 51 and 112.

required which decreases the annualized cost advantage of the injection system. Depending on site specific factors and reagent costs, levelized costs are equal for wet and dry injection FGD systems between one to two percent sulfur coal.

Several alkaline sorbents have been tested including lime, sodium carbonate, trona and nahcolite. Nahcolite is a naturally occuring mineral containing about 80 percent sodium bicarbonate, which has been found to be more reactive than sodium carbonate. The former loses two moles of $CO_2$ and one mole of water during reaction with $SO_2$, while sodium carbonate loses only one mole of $CO_2$ as shown below:

$$2NaHCO_3(s) + SO_2(g) \rightarrow Na_2SO_3(s) + H_2O(g) + 2CO_2(g)$$
$$Na_2CO_3(s) + SO_2 \rightarrow Na_2SO_3(s) + CO_2(g)$$

A typical nahcolite dry injection process is shown in Figure 17. As shown, solids formed in the first reaction are removed in a baghouse. Some $NaHCO_3(s)$ solids which are unreacted are also collected on the bag surface. These solids can continue to react with the $SO_2$ in the flue gas. The continued reaction of the unspent sorbent on the bag surface is one reason better results are obtained with baghouse collectors versus ESPs (113). Also, since a major portion of the $SO_2$ removal seems to occur on the bag surface, several methods of feeding have been tested. These include continuous, batch, and semi-batch (114).

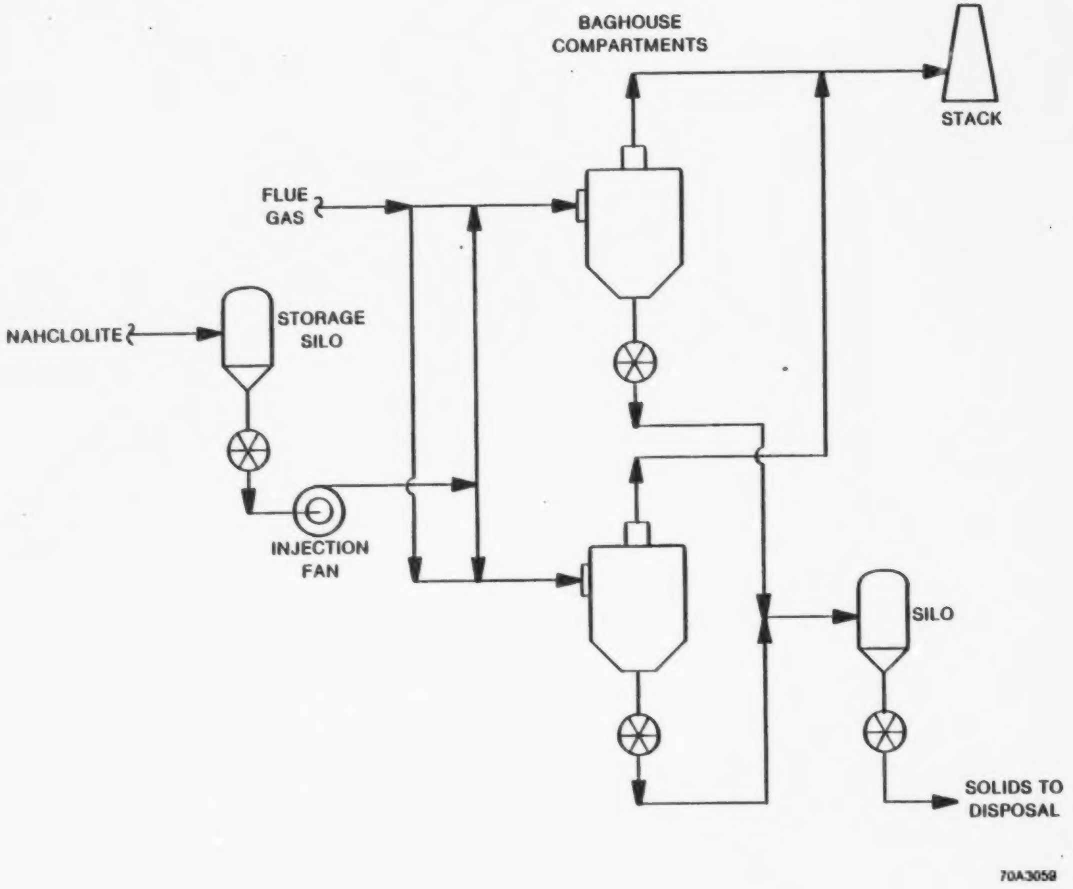

BAGHOUSE
COMPARTMENTS
STACK
FLUE
GAS
NAHCLOLITE
STORAGE
SILO
INJECTION
FAN
SILO
SOLIDS TO
DISPOSAL
70A3059

Figure 17. Nacholite Injection Process Flow Diagram

The $SO_2$ removal is a function of the feed ratio, the molar flow of absorbent divided by the molar $SO_2$ flow rate. Feed ratios of 1.5 to 2 are required to obtain high $SO_2$ removal. Other factors, such as gas-solids distribution and contact time are important in obtaining good removal.

Spent absorbent and reaction products can be partially recycled, as with spray drying systems to increase the reagent utilization. However, a ratio of two parts used to one part fresh reagent is generally the maximum practical ratio.

The desulfurization reactions using nahcolite as the sorbent in a dry injection system average only 70 percent completion, resulting in a dry waste containing $NaHCO_3$, $Na_2CO_3$, and $NaSO_4$ as well as fly ash. Disposal of untreated waste could be environmentally damaging because of the high levels of soluble sodium salts. Various treatment processes (pelletizing and/or chemical/physical stabilization) would probably be necessary. Landfilling in clay isolation cells has also been proposed, although in arid regions diked landfills may be adequate. Regeneration of sodium compounds from dry FGD waste using current technology from other fields is also being tested.

In summary, dry injection development programs are continuing under government and commercial funding. Greater $SO_2$ removal efficiencies have been demonstrated than were seen in earlier tests, but commercialization of this process has not occurred due primarily to sorbent cost and availability, relatively high feed ratios, and waste disposal concerns.

## Electron Beam

Development efforts are underway on an electron beam (E-beam) process for removing $SO_2$ and $NO_x$ from flue gas. Ebara Corporation of Tokyo began development work in 1970 and operated a 3MWe pilot unit in 1977-78 at a steel sintering plant. Avco Research Laboratories has conducted bench scale evaluations of the Avco/Ebara system since 1977 (115). Research-Cottrell is also involved in electron beam process development and has performed both bench scale testing and application analysis with the assistance of Ion

Physics Co. This work was carried out under a DOE contract in 1979-1980. Since then, Research-Cottrell has been awarded a contract to perform 10 MW pilot plant test at TVA's Shawnee test facility.

Both of these versions of the E-beam process are currently in the pilot or prototype development stage and are not, at present, ready for commercial application as coal fired units . A brief description of each version of the E-beam processes is presented below.

The electron beam process is based on irradiation of the flue gas to remove $SO_2$ and $NO_x$. As discussed above, two groups are currently working on electron beam processes. Both systems use an irradiation reactor to produce $SO_2$ and $NO_x$ containing solids. These solids are then removed from the flue gas in downstream particulate control devices.

In the Avco/Ebara process, ammonia is injected into the flue gas downstream of the primary particulate control ESP (See Figure 18). The ammonia then reacts in the reactor with the $SO_2$ and $NO_x$ to produce ammonium sulfate and nitrate. The reactions that take place are summarized in Figure 19. Ammonium sulfate and nitrate are removed from the flue gas downstream of the reactor in a second ESP. Pilot work performed by Ebara indicates that $SO_2$ removals of 90 percent and $NO_x$ removals of 80 percent are possible with this process (115).

A second configuration, proposed by Research-Cottrell, is shown in Figure 20 (118). This system uses a spray dryer for initial $SO_2$ removal. No ammonia injection is used. The E-beam reactor then follows the spray dryer to increase the $SO_2$ removal to 90 percent and remove 90 percent of the $NO_x$. The solid reaction products are removed in a baghouse downstream. Other configurations examined by Research-Cottrell include: 1) an electron beam/ESP combination with ammonia addition to produce potential fertilizer product, and 2) a dual alkali process which uses limestone to regenerate ammonia and produces gypsum for disposal or sale.

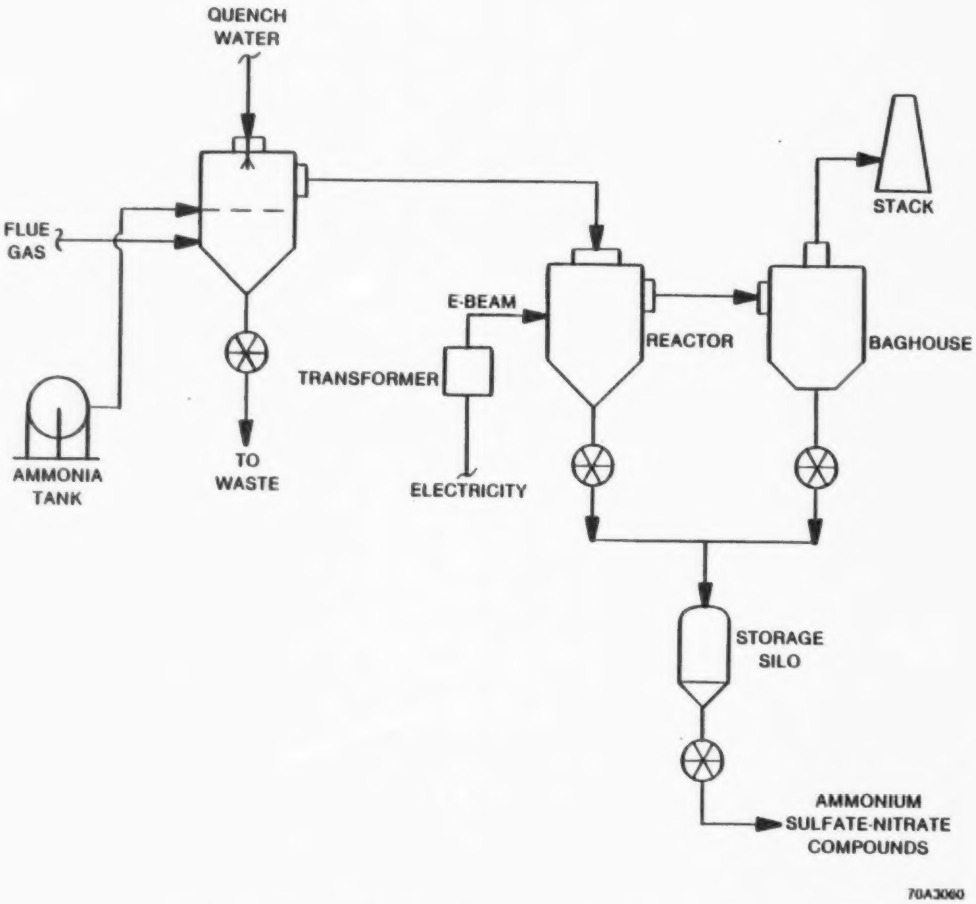

QUENCH WATER
FLUE GAS
AMMONIA TANK
TO WASTE
E-BEAM
TRANSFORMER
ELECTRICITY
REACTOR
STACK
BAGHOUSE
STORAGE SILO
AMMONIUM SULFATE-NITRATE COMPOUNDS
70A3060

Figure 18. AVCO/EBARA E-Beam Process Flow Diagram

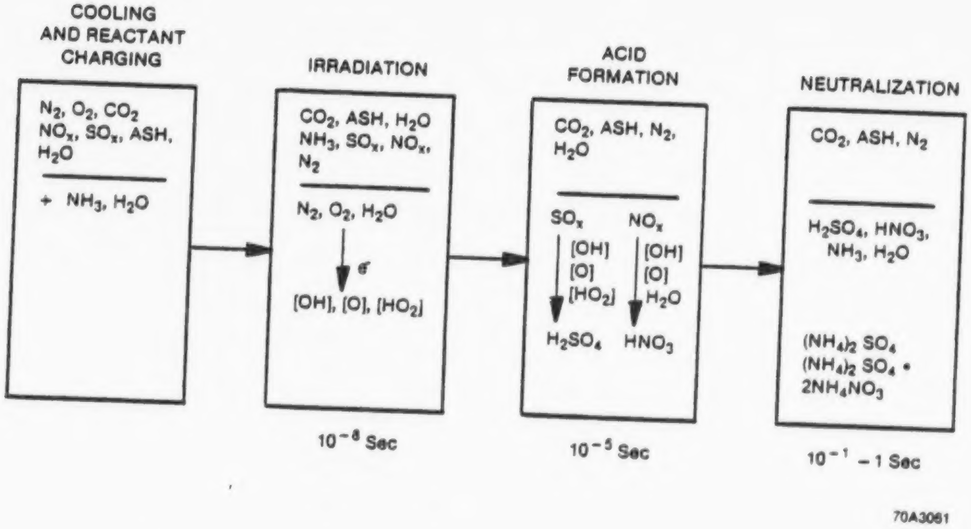
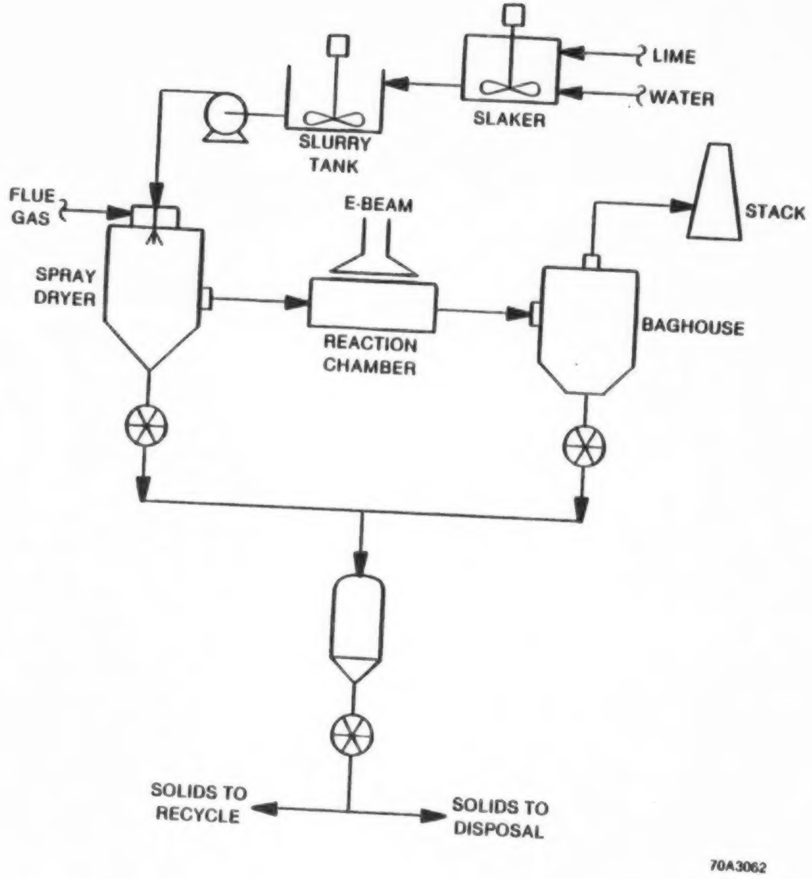

COOLING AND REACTANT CHARGING
$N_2$, $O_2$, $CO_2$ $NO_x$, $SO_x$, ASH, $H_2O$
+ $NH_3$, $H_2O$
IRRADIATION
$CO_2$, ASH, $H_2O$ $NH_3$, $SO_x$, $NO_x$, $N_2$
$N_2$, $O_2$, $H_2O$
$e^-$
[OH], [O], [$HO_2$]
$10^{-8}$ Sec
ACID FORMATION
$CO_2$, ASH, $N_2$, $H_2O$
$SO_x$
[OH] [O] [$HO_2$]
$H_2SO_4$
$NO_x$
[OH] [O] $H_2O$
$HNO_3$
$10^{-5}$ Sec
NEUTRALIZATION
$CO_2$, ASH, $N_2$
$H_2SO_4$, $HNO_3$, $NH_3$, $H_2O$
$(NH_4)_2SO_4$
$(NH_4)_2SO_4 \cdot 2NH_4NO_3$
$10^{-1} - 1$ Sec
70A3061

Figure 19. Generalized Reaction Chemistry of Electron Irradiation with Ammonia Injection



LIME
WATER
SLAKER
SLURRY TANK
FLUE GAS
E-BEAM
STACK
SPRAY DRYER
REACTION CHAMBER
BAGHOUSE
SOLIDS TO RECYCLE
SOLIDS TO DISPOSAL
70A3062

Figure 20. Research-Cottrell Spray Dryer/E-Beam System

## Shell Copper Oxide

The Shell Copper oxide process was developed by Shell Internationale Petroleum Maatschappij in the Netherlands and is licensed in the U.S. by UOP's Air Correction Division. This process uses a fixed bed of copper oxide supported an alumina for simulataneous $SO_2$ and $NO_x$ removal. Regeneration of the spent copper sulfate is accomplished by introducting a hydrogen containing flue gas which reduces the $CuSO_4$ to CuO (119).

To date, a completely integrated unit has not been applied to a coal-fired utility boiler. A 0.6 MW equivalent unit was tested at Tampa Electric Company's Big Bend Station beginning in 1974. Simultaneous removal of $SO_2$ and $NO_x$ was also tested at Big Bend during an EPA sponsored demonstration program in 1979-1980 (120). The largest commercial unit began operation in 1973 at Showa Yokkaichi Sekiyu (SYS) Refinery in Japan treating flue gas from a 40 MW equivalent oil-fired boiler. However, only the reactors and $SO_2$ gathering systems were installed due to the ready availability of refinery hydrogen and excess Claus plant capacity. (93,118,119).

A sample flow-scheme for the Shell Copper Oxide Process is shown in Figure 21. This scheme is based upon the SYS Refinery system.

$SO_2$ removal occurs in the reactor at approximately 400°C as shown below. Copper oxide supported on alumina reacts with $SO_2$ and oxygen in the flue gas to form copper sulfate.

$$CuO(s) + 1/2O_2(g) + SO_2(g) \rightarrow CuSO_4(s)$$

Copper sulfate produced during the $SO_2$ acceptance reaction shown above and, to a lesser extent, copper oxide also act as a catalyst for the reduction of $NO_x$ by injection $NH_3$ as shown below (120).

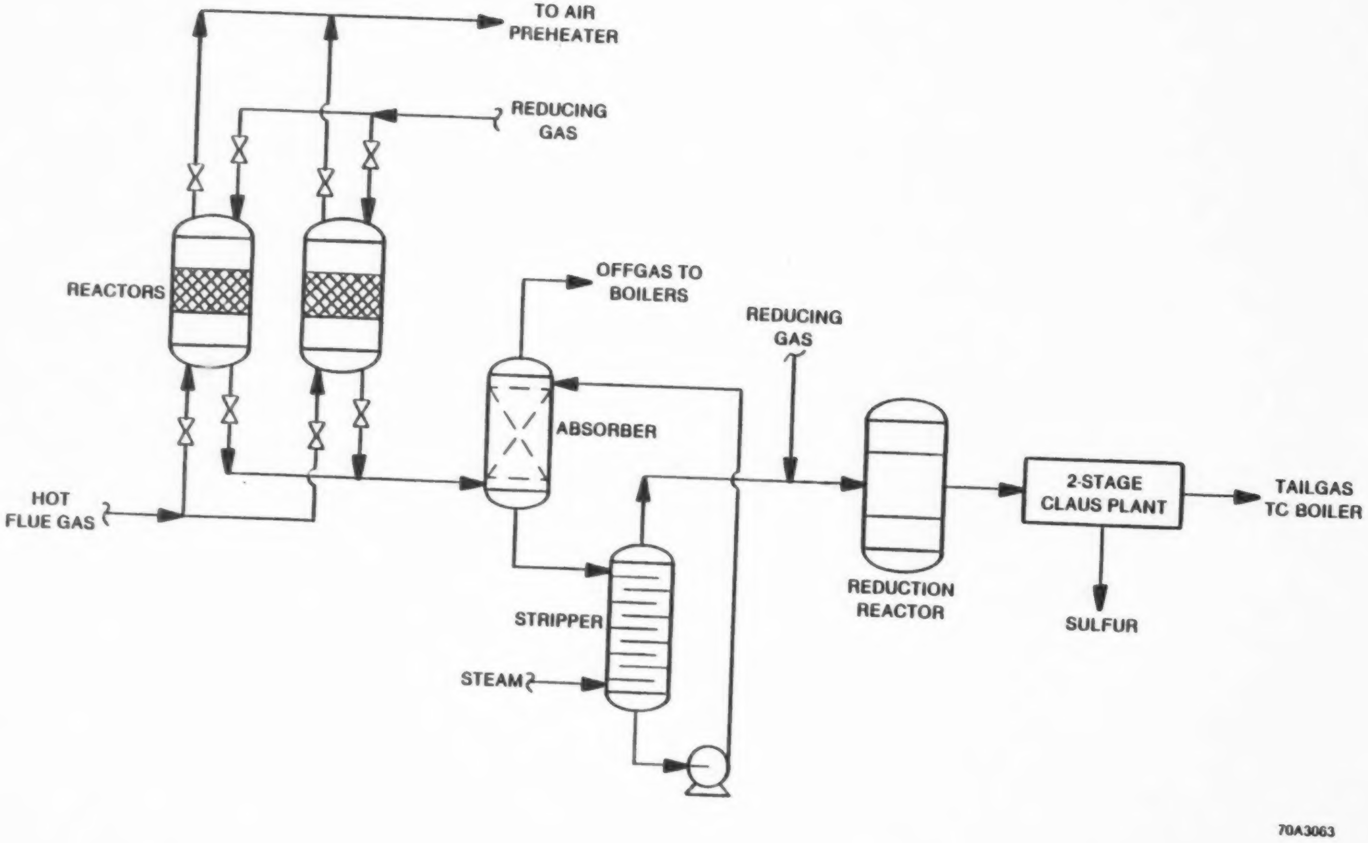

TO AIR PREHEATER
REDUCING GAS
REACTORS
OFFGAS TO BOILERS
REDUCING GAS
ABSORBER
HOT FLUE GAS
2-STAGE CLAUS PLANT
TAILGAS TC BOILER
STRIPPER
REDUCTION REACTOR
SULFUR
STEAM
70A3063

Figure 21. Shell Copper Oxide Process Flow Diagram

$$4NO(g) + 4NH_3(g) + O_2(g) \rightleftarrows 4H_2(g) + 6H_2O(g)$$

$$2NO_2(g) + 4NH_3(g) + O_2(g) \rightleftarrows 3N_2(g) + 6H_2O(g)$$

As more $SO_2$ is absorbed, the CuO available for reaction gradually becomes depleted. When the $SO_2$ emissions approach the design level, the reactor is shut down and isolated for regeneration of the copper absorbent. Due to the cyclic nature of the absorption and regeneration steps, spare reactors must be included to accept flue gas while spent sorbent is being regenerated.

Regeneration is accomplished by passing a reducing gas, hydrogen or methane, over the absorption bed. Using methane, regneration occurs by:

$$CuSO_4(s) + 1/2CH_4(g) \rightarrow Cu(s) + SO_2(g) + 1/2CO_2(g) + H_2O(g)$$

Simularly, regeneration with hydrogen is accomplished by:

$$CuSO_4(s) + 2H_2(g) \rightarrow Cu(s) + SO_2(g) + 2H_2O(g)$$

Oxygen in the flue gas oxidizes copper back to copper oxide at the beginning of the absorption or acceptance cycle. In general, the cycle time is between 40 and 80 minutes.

Several options are available for concentrating and preparing the $SO_2$ rich regeneration gases for further processing. These include water absorption/stripping solvent absorption, compression/condensation of water vapor, and complete liquefaction. (93,119,121, 122). The concentrated $SO_2$ stream can be processed to produce sulfuric acid, liquid $SO_2$, or elemental sulfur. Figure 21 shows, as an example, the production of elemental sulfur using a two stage Claus unit.

# SECTION 3

## CONTINUOUS EMISSIONS MONITORING

In this section, the $SO_2$ removal performance of various FGD systems is presented based on continuous emissions monitoring results. The current NSPS requires a percentage $SO_2$ reduction as well as a limit on $SO_2$ emissions. For coal-fired boilers, flue gas desulfurization is required to meet the $SO_2$ emission reduction and limitation requirements. The NSPS also requires continuous monitoring of $SO_2$ emissions both into and out of the FGD system (5).

In May 1977, EPA began a program to acquire $SO_2$ monitoring data in support of possible revisions to the initial utility boiler NSPS (18). The NSPS, originally promulgated in December 1971, was ultimately revised and promelgated on June 11, 1979. The continuous monitoring data for FGD $SO_2$ removal efficiencies that EPA obtained are summarized in this section. These data were obtained from continuous monitoring of nine utility FGD systems.

Additionally, continuous emissions monitoring data from four other utility FGD systems and one FGD system applied to an industrial boiler are included. These data were obtained after the promulgation of the revised NSPS.

Table 57 summarizes the mean removal efficiencies and standard deviations achieved by the 14 FGD systems. The remainder of this section provides, for each of the 14, more information on the plant and FGD system, monitoring results and any problems encountered, and monitoring instrumentation.

The data represent results from individual ealuations and are extracted from final site reports. In four instances; A. B. Brown Unit No. 1, Southwest Unit No. 1, Pleasants Unit No. 1, and Cane Run Unit No. 6; the data are undergoing rigorous analysis by OAQPS and these results may therefore differ.

TABLE 57. SUMMARY OF FGD SYSTEM PERFORMANCE

| Plant Information | | | FGD System Information | | |
|---|---|---|---|---|---|
| Name and Location | Size (MW) | Coal Sulfur Content (%) | Type of System | Efficiency (%$_{\mu \pm \sigma}$) | Emission Rate (lb/SO$_2$/10$^6$ Btu) |
| Tennessee Valley Authority,<br>Shawnee Prototype,<br>Paducah, Kentucky | 10[a] | 3.0 | Limestone and Adipic Acid, Venturi-Spray Tower | 96.1 ± 1.2 | 0.22 |
| Kansas City Power and Light,<br>Lawrence Unit No. 4,<br>Lawrence, Kansas | 125 | 0.5 | Limestone, Rod Scrubber/ Spray Tower (2 modules) | 96.6 ± 5.3 | 0.03 |
| Kansas City Power and Light<br>LaCygne Unit No. 1<br>LaCygne, Kansas | 820 | 5-6 | Limestone, Venturi/Tray Absorber (8 modules) | No data[b] | - |
| Springfield City Utilities,<br>Southwest Power Plant<br>Unit No. 1<br>Springfield, Missouri | 194 | 3.5-4.0 | Limestone and Adipic Acid or DBA, TCA (1 module), Tray Absorber (1 module) | >90[c] | <0.64 |
| Louisville Gas Electric,<br>Cane Run Unit No. 4<br>Louisville, Kentucky | 182 | 4.1 | Carbide lime, Packed Absorber (2 modules) | 83.8 ± 4.3 | 0.88 |
| Columbus and Southern<br>Ohio Electric<br>Conesville Unit No. 5<br>Conesville, Ohio | 375 | 3.5-4.5 | Thiosorbic lime, Packed Absorber (2 modules) | 89.2 ± 3.6 | 0.80 |
| Pennsylvania Power<br>Bruce Mansfield Unit No. 1<br>Pittsburgh, Pennsylvania | 917 | 3.0 | Lime, Two-stage Venturi Absorber (6 modules) | 80.6 ± 4.6,<br>85.3 ± 3.4 | 1.01,<br>0.77 |
| Tennessee Valley Authority<br>Shawnee (TCA) Prototype<br>Paducah, Kentucky | 10[a] | 3.0 | Lime, Packed Absorber | 88.6 ± 2.2 | 0.64 |
| Allegheny Power Systems,<br>Pleasants Unit No. 1<br>Willow Island, West Virginia | 640 | 3.0 | Lime, Venturi Spray Tower, (4 modules) | >90 | <0.50 |
| Southern Indiana Gas and Electric<br>A. B. Brown Unit No. 1,<br>Evansville, Indiana | 265 | 3.5 | Dual Alkali, Impingement Tower (2 modules) | ~87 ± 4 | 0.85 |
| Louisville Gas and Electric,<br>Cane Run Unit No. 6<br>Louisville, Kentucky | 300 | 4.8 | Dual Alkali, Tray Tower Absorber (2 modules) | 91.6 | 0.70 |
| Northern Indiana Public Service,<br>D. H. Mitchell Unit No. 11<br>Gary, Indiana | 115 | 3.2 | Wellman-Lord, Three-stage Tray Absorber | 89.6 ± 1.0 | 0.61 |
| Philadelphia Electric,<br>Eddystone Unit No. 1A,<br>Philadelphia, Pennsylvania | 325 | 2.6 | Magnesium Oxide, Venturi Scrubber (3 modules) | 96.8 ± 1.2 | 0.16 |
| Celanese Amcelle,<br>(Industrial)<br>Cumberland, Maryland | ~45 | 1.9 | Lime, Spray Dryer | 70 ± 9 | 0.92 |

[a] The boiler is actually rated at 175 MW, but the FGD system treats flue gas equivalent to 10 MW.

[b] Problems experienced with the O$_2$ monitoring system resulted in the discontinuation of monitoring before any useful data were accumulated.

[c] Variety of operating conditions examined.

## LIMESTONE SYSTEMS

Continuous emission monitoring sponsored by the EPA has been performed at four limestone FGD systems. These units are the Shawnee venturi-spray tower prototype, Lawrence 4, LaCygne 1, and Southwest 1. Data was not obtained at LaCygne due to boiler and process operating problems. The four units tested are somewhat atypical compared to the design of current limestone units. The Shawnee unit is a 10 MW prototype FGD system which has been used for many years to test lime and limestone process variations. The Lawrence 4 unit is a 125 MW plant burning low sulfur coal. A rod and spray tower is used for $SO_2$ control. The LaCygne unit is one of the oldest commerical scale FGD systems in operation and burns one of the least desirable coals in the country. The venturi-tray tower arrangement used there, however, is not a popular absorber. Finally, the Southwest 1 unit of City Utilities of Springfield initially used turbulent contact absorbers. The absorbers have since been modified to tray towers. Further details about each plant and the results obtained are presented below. During the next review of the utility boiler NSPS, it is most probable that data from several spray tower systems will be added to the limestone data base.

### Shawnee (Venturi/Spray Tower)

#### Plant and FGD System Description--

The Shawnee Power Station, operated by the Tennessee Valley Authority (TVA), is a coal-fired steam generating station having 10 turbines, each served by a boiler and stack. Three percent sulfur bituminous coal is burned. Unit No. 10 is rated at 175 MW. A portion of the exhaust gases from Unit No. 10 is directed for treatment by one of two prototype scale wet scrubber systems: a venturi with a spray tower, and a turbulent contact absorber (TCA). Each pilot FGD scrubber treats the equivalent of up to 10 MW of flue gas (~30,000 acfm). This section describes testing performed on the venturi/spray tower (V/ST) system (123). Lime testing on the TCA system is described in another section.

In the V/ST system, flue gas at about 1500 to 4500 ppm $SO_2$ first enters the venturi scrubber, which removes the bulk of the particulate matter as well as some $SO_2$. The venturi is adjustable to permit control of pressure drop. From the venturi, the flue gas passes through a spray tower for further $SO_2$ removal, is reheated, and exits through the stack.

Scrubber liquor is collected in a holding tank located below the venturi and spray tower. Limestone and any enhancement agents, such as adipic acid, are added to this holding tank.

Continuous Emissions Monitoring Test Results—

FGD system performance monitoring has been ongoing at the Shawnee test facility since late 1976. The effects of varying many system parameters have been reported (123, 124, 125, 126, 127). The limestone venturi-spray tower system has been shown capable of achieving better than 90 percent $SO_2$ removal under certain operating conditions.

Continuous emission monitoring began in January 1979. During testing, adipic acid was added to the holding tank and maintained at 1500 ppm. Thirty-one full calendar data days (18 hours or more of monitoring data) were obtained. Over those 31 days, the average $SO_2$ removal efficiency was 96.1 percent, and the standard deviation was 1.2 percent (128).

Continuous Emission Monitoring Systems—

$SO_2$ and $O_2$ concentrations were measured upstream and downstream of the FGD system using continuous emission monitors. The $SO_2$ instrument at both locations was a DuPont 460 analyzer. Thermox Model WDG 3 instruments were used to measure $O_2$. Measurements were recorded continuously on strip chart recorders.

A moisture content correction for the $SO_2$ measurements was determined from the temperature measurement in the sample gas stream. The gas was assumed saturated at this temperature. The moisture content correction for the $O_2$ gas measurements was the stack moisture content as the $O_2$ analyzers measured a wet gas stream.

All the continuous emission monitoring instrumentation described above was subjected to the appropriate test procedures specified in the Federal Register (Vol. 40, No. 194, 1975) to assure the accuracy of measured data (128).

## Lawrence Unit No. 4

### Plant and FGD System Description--

The Lawrence Energy Center, a power generating station with a capacity of 625 MW (gross), is owned and operated by Kansas Power and Light. The station consists of five power generating units. In service since 1959, it currently operates as a cyclic-load unit.

The 125 MW steam generator for Lawrence 4 is a balanced-draft, tangential-fired, multiple-fuel-burning unit. Although it was designed to burn pulverized coal, oil and/or gas in any combination, it is now fueled exclusively by a low-sulfur subbituminous grade of coal which originates from mines located in Medicine Bow in the southeast section of Wyoming. This coal contains on the average 0.5 percent sulfur, 10 percent ash, and 12 percent moisture, and has a heating value of 10,000 Btu/lb. At full load, Lawrence 4 consumes approximately 50 tons of coal per hour.

To meet air emission regulations of the Department of Health and Environment of the State of Kansas and the U.S. EPA, Unit No. 4 is equipped with a tail-end wet limestone scrubbing system consisting of two parallel scrubber modules for the control of particulate and sulfur dioxide. This system represents a second generation design replacement of the limestone furnace-injection systems originally installed on the unit in 1968.

Flue gas at 280°F enters the scrubbing system at a rate of 403,000 acfm through two parallel, rectangular, rod-deck venturi scrubbers, each comprised of a converging gas section and rod section. The converging gas section directs the flue gas downward to the rod-deck section and consists of two staggered levels of rubber-coated fiberglass rods. The vertical spacing between the two rows of rods is automatically controlled according to gas load in order to insure a constant gas-side pressure drop across the rod section.

A series of nonatomizing, fan-type spray nozzles located around the perimeter of the throat area continuously spray limestone slurry into the rod-deck scrubber, where an intimate gas-slurry contact occurs, which facilitates particulate and sulfur dioxide removal. Spent slurry from the rod section gravity feeds into a collection tank located directly below the venturi. This tank, which has a liquid capacity of 50,100 gal, retains the slurry for approximately 14 minutes to allow for completion of chemical reactions. The slurry is recycled from the collection tank to the rod-deck scrubber by means of a slurry recirculation pump (one operational/one spare) at a rate of 3600 gpm.

After passing through the rod-deck venturi, the flue gas makes one 90 degree turn as it approaches the spray tower and another 90 degree turn before passing upward through the spray tower at 349,000 acfm. The saturated gas, cooled to 124°F, flows upward through two levels of sprays in the open towers, where the gas is contacted by the slurry, which is sprayed counter-current to the gas flow. The spray levels, each of which include four internal spray headers containing six spray nozzles each, are situated at approximately 10 foot intervals above the inlet duct.

Spent slurry from each spray tower gravity feeds into a reaction tank located directly below the tower. The reaction tank, which has a liquid capacity of 69,200 gal, retains the slurry for approximately 10 minutes for completion of chemical reactions and dissolution of fresh limestone additive. The slurry is then recycled to the spray tower by means of one slurry recirculation pump at a rate of 5300 gpm.

Entrained droplets of moisture and slurry picked up by the flue gas stream are removed in a mist elimination section approximately 10 feet above the spray zone in the spray towers. Each mist eliminator is an A-frame design comprised of a bulk entrainment separator followed by two stages of chevron vanes. Each is equipped with an intermittent, high-pressure, waterwash system which sprays blended makeup water on the top of the bulk entrainment separator and the bottom of the first mist eliminator stage.

Following passage through the mist eliminators, the saturated gas stream is reheated by an in-line, carbon steel reheater. One such reheater, which consists of four rows of circumferential finned tubes arranged in a staggered fashion, is provided for each spray tower. The heating medium is hot water from the boiler feed water deaerator. Two half-track soot blowers located upstream of the reheaters provide compressed air every 4 hours of service to keep the reheaters clean. The reheaters boost the temperature of the gas stream approximately 20°F, after which it flows through the ducts leading to the induced-draft fans and stacks, which discharge it to the atmosphere (129).

### Continuous Emissions Monitoring Test Results--

Continuous emission monitoring of the limestone FGD system began in January 1979. Twenty-two full calendar data days (18 hours or more of monitoring data) were obtained. The average $SO_2$ removal efficiency over that time period was 96.6 percent, with a standard deviation of 5.3 percent. The boiler load during CEM testing averaged 85 MW (128).

### Continuous Emission Monitoring Systems--

The continuing emission monitoring system used at Lawrence No. 4 consisted of an extraction/conditioning module and a mobile laboratory that housed the analyzers. Thermo Electron Model 40 pulsed fluorescence analyzers were used to monitor $SO_2$ at both the scrubber inlet and outlet. A Beckman

Model 742 polarographic analyzer was used to monitor the inlet $O_2$ concentration while an MSA Model 802 paramagnetic analyzer was used to monitor the outlet $O_2$ concentration. Measurements were recorded continuously on strip chart recorders.

All the continuous emission monitoring instrumentation described above was subjected to the appropriate test procedures specified in the Federal Register to assure the accuracy of measured data (128).

## LaCygne Unit No. 1

### Plant and FGD System Description--

The 820 MW LaCygne No. 1 Unit began commercial operation on June 1, 1973 as a joint project of Kansas Gas and Electric Company and Kansas City Power and Light Company. The companies share equally in ownership and output and the unit is operated by KCP&L. Construction of Unit No. 1 began in 1969, and erection of the air quality control system was initiated in mid-1971.

Unit No. 1 fires a low grade, subbituminous coal with an as-fired heating value of 9,000 to 9,700 Btu/lb, and an ash content of 25 percent and sulfur content of 5 to 6 percent.

The boiler for Unit No. 1 is a cyclone-fired, supercritical, once-through, balanced-draft unit with a rating of 6,200,000 pounds of steam per hour, 1,010 degrees F, 3,825 psig at the superheat outlet. The turbine-generator is rated at 874 MW gross output with five percent overpressure and 3,500 psi throttle pressure. Three auxiliary, oil-fired boilers are used for plant startup or for powering a 20 megawatt house turbine-generator. The net plant output is 820 megawatts, adjusted to include 24 megawatts used by the AQC system and 30 megawatts by plant auxiliaries.

The air quality control system consists of eight two-stage venturi-absorber scrubber modules designed to treat the boiler flue qas flow of 2,760,000 ACFM. (345,000 ACFM per module at 285 degrees F.) The duct work design does not provide for flue gas bypass of the system, but each module can be isolated for maintenance by individual dampers. On-site limestone grinding and slurry storage facilities provide up to 1,000 tons of slurry per hour. The unit has a balanced draft system with three 7,000 hp forced draft fans and six 7,000 hp induced draft fans located between the air quality control system and the 700 foot stack. There is a common plenum at both the scrubber inlet and outlet. Spent slurry and fly ash are removed from the module recirculation tank through rubber-lined pipes to the settling pond at the rate of 3,500 tons of solids per day. Clear makeup water is pumped from the pond and the loop is closed by recycling ball mill and module makeup water back into the system.

As the hot flue gas enters the venturi, it is sprayed with slurry from 48 spray and 32 wall wash nozzles resulting in up to 99 percent of the particulate removal. The gas then makes a 180 degree turn up through the absorber section. In the reaction chamber, the $SO_2$ is removed as the gas is forced through a limestone slurry solution sprayed on stainless steel sieve trays. The cleaned gas passes through mist eliminators to remove moisture and then is reheated to avoid deposits on the fans and provide buoyancy from the stack (130).

Continuous Emissions Monitoring Test Results—

Continuous emission monitoring tests at LaCygne Unit No. 1 was part of an EPA program to acquire data in support of possible NSPS revisions. Data gathering began in the late summer of 1977. Unfortunately, continuous emission monitoring testing was discontinued at LaCygne in October 1977 with little useful data accumulated.

After several failures with a single extraction line for both $SO_2$ and oxygen, it was decided that a separate probe and sample line system would be necessary. This system was not installed due to test termination at this facility.

A single four hour full load test was completed on August 26, 1977. $SO_2$ removal efficiency averaged 77 percent with individual modules achieving 65 to 80 percent removal (130).

Continuous Emission Monitoring Systems—

The FGD inlet $SO_2$ instrument was a single-point, dual-range DuPont Model 460 analyzer. It operated on a 10-minute cycle and recorded on a continuous trace chart recorder.

To monitor inlet $O_2$, a Beckman Model F3 paramagnetic analyzer was originally installed on the $SO_2$ sample stream exhaust. However, this arrangement did not function reliably due to interface problems. The system was modified to provide a separate probe and sample line for the $O_2$ analyzer. This arrangement operated satisfactorily.

The FGD outlet $SO_2$ monitor was a DuPont Model 460 two-point, dual-range analyzer. Like the inlet monitor, it operated on a 10 minute cycle and recorded results on a continuous trace recorder (131). As mentioned earlier, $O_2$ was never properly monitored at the outlet.

## Springfield City Utilities

### Plant and FGD System Description—

Springfield City Utilities' Southwest Power Plant Unit 1 is a 194 MW unit designed to burn 3.5 to 4.0 percent sulfur Western Kentucky bituminous coal. Flue gas from the boiler passes through an air preheater, an electrostatic precipitator, and induced draft fans prior to entering the limestone FGD system and exiting the plant through the stack.

The limestone FGD system for Unit 1, as originally designed, consisted of two parallel turbulent contact absorber (TCA) modules. Each was sized to handle 60 percent of the design flue gas flow. One of these modules (module S-1) was converted to a tray absorber in May 1981 (during the period of documented testing and emissions monitoring). The other (module S-2) was converted in October 1981 (after the completion of testing and emissions monitoring). Limestone preparation and sludge handling systems are common to the two FGD modules.

As the FGD system normally operates, flue gas, typically 2000-2600 ppm $SO_2$, enters a presaturator and is cooled from 300 to 125°F by supernatant liquor (overflow from the sludge thickener). The gas then passes through three FGD contactor levels where $SO_2$ is removed. From there, the gas goes through a mist eliminator and out the stack. The flue gas flow rate through each scrubber module is approximately 300,000 acfm (saturated at 120°F) at full load conditions.

Scrubber feed slurry is circulated from a reaction tank below the scrubber through spray nozzles above the top level of TCA spheres (now trays) at a nominal rate of 13,000 gpm (L/G ratio of 40-45 gal/acf at full load conditions). Makeup limestone is added from a limestone storage tank to the reaction tank to maintain the desired pH. The solids content of the recirculating slurry is controlled by pumping slurry to the thickener. Solids are first settled in the thickener, and then the thickener underflow slurry is further dewatered on a vacuum filter belt. The resulting sludge is mixed with fly ash in a pug mill and trucked to an on-site landfill. Supernatant (thickener overflow) from the dewatering operation is returned to the FGD modules as presaturator and mist eliminator wash water makeup. Additional supernatant water is also utilized in the limestone preparation area (46).

### Continuous Emissions Monitoring Test Results—

In January 1981, a test program sponsored by the Industrial Environmental Research Laboratory of the EPA was initiated at Southwest Power Plant. Testing continued through October 1981. The main objective was to demonstrate the effectiveness of adipic acid as an $SO_2$ removal enhancement agent for limestone FGD systems. A second objective was to validate the Shawnee data base, thereby demonstrating its utility as a tool in designing full-scale FGD systems.

Two relatively simple process modifications were made for the test program; an adipic acid feed system and a forced oxidation system were installed. Both systems were designed for temporary use during the test program. The adipic feed system consisted of a two-ton vibrating silo, a screw conveyor, and a vibratory screw feeder. The adipic acid was fed to the limestone sump, mixed with freshly ground limestone, and then stored in the limestone storage tank. Since the limestone handling equipment is common to both scrubber modules, both scrubbers usually operated with about the same adipic acid concentration.

The forced oxidation system included four 1800 scfm compressors that discharged into a common header. Air flow to each scrubber module could be controlled independently. The air sparger consisted of 60 one-inch holes cut in a six-inch pipe which was located along the perimeter of the reaction tank. This design limited the sheer stress exerted by the air on the agitator, but the agitator horsepower was not used to help distribute the air as might be recommended in a permanent installation. The expense of upgrading the agitator assembly could not be justified for the short-term testing program.

Testing with adipic acid was performed under two modes: the first using forced oxidation, the second using natural oxidation. Further testing was conducted using an alternative source of buffer, dibasic acid (DBA). Table 58 summarizes the $SO_2$ removal results of this testing. The boiler load during testing varied as required to meet load.

The data presented in Table 58 demonstrate the significant improvements in $SO_2$ removal achieved as a result of using adipic acid and DBA. Baseline $SO_2$ removal efficiencies (those where no adipic acid or DBA were used) ranged from 46 to 72 percent. With the addition of adipic acid or DBA, $SO_2$ removal efficiencies were generally near or above 90 percent.

Several observations of FGD system operation and performance were noted during the testing period. First, scrubber reliability (hours operated/ hours required x 100) improved over the course of the test program. Reliabilities for both modules were about 85 percent, an improvement over reliabilities of

## TABLE 58. SUMMARY OF TEST RESULTS AT SWPP

| Mode | Module | Operating Time (days) | Scrubber Feed pH | Adipic Acid Concentration (ppm) | 24-Hour Average $SO_2$ Removal (%μ + σ) | Notes |
|---|---|---|---|---|---|---|
| Forced Oxidation | S-1 | 20 | 5.4 | 0 | 46 ± 8.3 | 1 |
|  | S-1 | 6 | 5.4 | 1340 | 84 ± 3.9 | 2 |
|  | S-2 | 24 | 5.4 | 0 | 58 ± 12.2 | 1 |
|  | S-2 | 49 | 5.0 | 2410 | 91 ± 6.5 | 3 |
|  | S-1 | 40 | 5.1 | 2340 | 89 ± 7.3 | 3 |
| Natural Oxidation | S-1 | 24 | 5.6 | 0 | 68 ± 4.5 | 1 |
|  | S-1 | 39 | 5.2-5.5 | 640-1000 | 88 ± 4.2 | 4 |
|  | S-1 | 17 | 5.4 | 1650 | 96 ± 1.9 | 5 |
|  | S-2 | 24 | 5.6 | 0 | 72 ± 6.1 | 1 |
|  | S-2 | 39 | 5.2-5.4 | 580-890 | 87 ± 6.1 | 6 |
|  | S-2 | 13 | 5.2 | 1590 | 94 ± 3.5 | 3 |
| Natural Oxidation with DBA Addition | S-1 | 16 | 5.4 | 480* | 90 ± 3.4 | 4 |
|  | S-1 | 13 | 5.4 | 1310* | 96 ± 1.9 | 5 |
|  | S-2 | 16 | 5.4 | 420* | 78 ± 9.5 | 6 |
|  | S-2 | 11 | 5.4 | 1020* | 94 ± 1.6 | 3 |

*DBA Concentration (ppm as adipic acid)

Key to Notes:
1 - Baseline testing (with no adipic acid added).
2 - Testing without TCA spheres. Goal: 85% $SO_2$ removal.
3 - Testing with TCA spheres. Goal: 95% $SO_2$ removal.
4 - Testing after conversion from TCA to tray tower. Goal: 85% $SO_2$ removal.
5 - Testing after conversion from TCA to tray tower. Goal: 95% $SO_2$ removal.
6 - Testing with TCA spheres. Goal: 85% Se removal.

Source: Reference 46

30-60 percent reported during 1979 and 1980. This improvement was probably a result of: 1) a number of mechanical changes implemented by the City Utility staff, 2) revised process operating procedures, 3) additional process engineering input, and 4) addition of adipic acid.

During the natural oxidation testing phase, the TCA module (S-2) was not maintained well since the unit was to be revamped at the conclusion of the test program. This resulted in buildup of solids in the TCA cages and some TCA sphere failure. These problems accounted for the lesser $SO_2$ removal efficiencies achieved by the TCA module (module S-2) during the natural oxidation with DBA addition as compared to the tray tower module. Periods of higher pH and lower limestone utilization during the natural oxidation phase also contributed to less favorable performance as opposed to forced oxidation (46).

### Continuous Emission Monitoring Systems--

Continuous emissions monitoring of the flue gas entering each of the two scrubber modules was handled by a DuPont 400 $SO_2$ analyzer modified for two-point sampling and a Thermox Model WDG III $O_2$ analyzer. For analysis of the flue gas exiting each scrubber, a DuPont Model 460 and Thermox WDG III analyzer arrangement was used. The signal outputs from the analyzers were recorded by a DART microprocessor. A chart recorder served as backup.

This entire monitoring system was tested according to the "Proposed Rules" published in the Federal Register (Vol. 44, No. 197) for certification of an emissions monitoring system. In all cases the continuous monitoring system at SWPP met the required criteria (46).

### Current Status--

In October 1981, module S-2 was converted from TCA spheres to trays. As noted earlier, module S-1 had similarly been converted in May of 1981. The use of these stainless trays has considerably improved the availability of both modules, although no improvement has been noted in $SO_2$ removal efficiency as a direct result of the conversions. The repacement of the spheres was justified economically by the fact that the trays installed had an economic payout of only one year when compared to the annual replacement and maintenance costs associated with the spheres (34).

As a result of the adipic acid/DBA testing program, City Utilities decided to permanently enhance the FGD system removal efficiency by adding DBA. In December 1981, a temporary DBA storage/feed system was put into service. (A permanent system is planned). The addition of DBA has improved in the performance of the FGD system. $SO_2$ concentrations at the FGD outlet have decreased, and the average emission rate for the first quarter of 1982 was 1.0 lbs/MMBtu as compared to the regulatory limit of 1.2 lbs/ MMBtu (35).

## LIME SYSTEMS

Five lime systems have been analyzed for $SO_2$ removal efficiency with continuous emission monitors by the EPA. These plants cover a wider range of sizes and sulfur levels than the three limestone plants. The units are Cane Run 4, Conesville 5, Bruce Mansfield 1, the Shawnee TCA, and Pleasants 1.

### Cane Run 4

### Plant and FGD System Description—

The Cane Run Power Station is owned and operated by Louisville Gas and Electric. It is comprised of six steam electric generators that are capable of producing a maximum of 1007 MW. Cane Run Unit No. 4, which has been in service for 19 years, is rated at 182 MW.

A high sulfur bituminous-grade Kentucky coal is burned at Cane Run. This coal has an average heating value of 11,000 Btu/lb, and average ash, moisture, and sulfur contents of 14.1, 9.6, and 4.1 percent, respectively.

Unit No. 4 is equipped with an electrostatic precipitator for the control of fly ash upstream of the FGD system. The commercial startup date for the carbide lime slurry FGD system was August 1977. The FGD system installed on Cane Run 4 consists of two parallel scrubber modules designed to treat 734,000 acfm of flue gas at 325°F when the unit is operating at full load. The FGD system includes gas handling and treating equipment, slurry handling equipment, solids concentrating equipment, waste disposal and pond water return equipment, and lime preparation and handling equipment.

The flue gas exits the boiler and passes through existing ESPs at 734,000 acfm and 325°F. Flue gas from existing induced-draft fans discharge into the FGD system. The ductwork and damper network provided with the FGD system allows gas to partially or totally bypass the scrubber modules. Guillotine isolation dampers installed at the inlet of each booster fan, at the outlet of each scrubber module, and in the bypass breeching enables gas to bypass one or both scrubber modules during boiler operation.

Following passage through the booster fans, the gas enters the scrubbing modules. Each scrubber module consists of a verticle absorber tower preceded by a quencher and flooded elbow. Each quencher is a wetted-wall conical frustrum section in the duct. A series of nozzles in the quencher inject lime scrubbing slurry into the gas stream to insure thorough wetting of the gases prior to paseage through the absorber. Immediately below each quencher is a flooded elbow. This section serves as a catch basin for the spent quencher slurry and complete the saturation of the gas stream. Some removal of sulfur dioxide occurs in the quencher and flooded elbow since part of the lime slurry recycle stream is diverted to these sections for wetting and saturation.

The quenched flue gas enters the base of each absorber tower at 291,500 acfm and 127°F. Each absorber tower is a single stage mobile bed contactor. The mobile bed contactor contains a fluid bed packing of solid spheres which serve to break up the gas stream and provide pockets for intimate mixing of the flue gases and scrubbing slurry. The flue gas passes upward through the packing where it contacts the scrubbing slurry sprayed into the gas stream countercurrently through large, low pressure, slurry sprays.

Entrained droplets of moisture and slurry picked up by the gas stream due to the turbulent mixing of slurry and gas in absorption zone are removed by mist eliminators. Each absorber is equipped with a two-stage, two-pass, chevron-type mist eliminator located in the top portion of each tower. Each mist eliminator is equipped with its own set of water sprays to retard the accumulation of solids which build up on the chevron blades.

Following passage through the mist eliminators, the cool, saturated gas stream is reheated by oil-fired burners located in the discharge ducts entering the stack. The direct oil-fired reheaters boost the temperature of the gas stream approximately 40 to 50°F prior to discharge to the existing stack.

Each scrubber module is equipped with its own compartmentalized reaction tank, recirculation pumps, and recirculation line for contacting the flue qas with scrubbing slurry. Three recirculation pumps deliver 17,625 gpm of 10 percent solids scrubbing slurry to each scrubber module. Of this amount, 1,760 gpm is provided to the quencher and 15,865 gpm is provided to the absorber. This slurry, as well as 80 gpm of mist eliminator wash water, drains to a cone-shaped reservoir located at the base of each absorber. The spent slurry and wash water then drains through a main pipeline to the return section of the reaction tank.

The effluent bleed pumps discharge the waste solids accumulated in the slurry loop to the thickener. Approximately 220 gpm of slurry is discharged from the feed section of each reaction tank. The thickener concentrates the waste solids from approximately 10 to 25 percent. In order to aid the thickening process, a polyelectrolyte feeding system is provided to enhance precipitation within the thickener.

Sludge is removed from the bottom of the thickener to an on-site pond for final disposal. Clarified overflow from the thickener gravity flows to the thickener overflow return tank for return to the reaction tank return sections. Supernatant from the sludge pond is added to the thickener overflow return tank to maintain system liquid levels (132).

### Continuous Emissions Monitoring Test Results—

Continuous emission monitoring was conducted at Cane Run Unit No. 4 from July throuqh December 1977. Based on 89 twenty-four hour periods, the FGD system achieved an $SO_2$ removal efficiency of 83.8 percent with a standard deviation of 4.3 percent. Boiler load generally varied between about 100 MW and capacity (131).

Among the problems encountered with the FGD system were excessive pressure drops due to gas flow distribution problems in the ducts and mist eliminators, and poor operation of guillotine dampers. Prior to the testing described above, redesign and modifications of the system were performed in order to increase $SO_2$ removal, improve mist eliminator efficiency, and correct a number of material failures with the coatings applied to the outlet ducts and stack.

### Continuous Emission Monitoring Systems—

$SO_2$ concentration at the inlet and outlet was continuously monitored by a single DuPont Model 460 four-point, dual range analyzer. The instrument operated on a 14-minute cycle. During a cycle, each of four measurement points was sampled (i.e., north module inlet, north outlet, south inlet, south outlet). Measurement results were printed on a multipoint chart recorder in the scrubber control room. A secondary single-pen continuous trace recorder was used as a backup.

A Scott Model 150 paramagnetic analyzer was used to measure $O_2$. Initially, the $O_2$ analyzer was connected directly to the $SO_2$ analyzer. The system was later modified to incorporate a condenser for moisture removal prior to $O_2$ measurement. $O_2$ measurements were recorded on a single-pen continuous trace recorder. For each 14-minute cycle, approximately 2-1/2 minutes of continuous record were available for each of the four sample points.

The moisture content of the sample stream was determined by manual techniques. Moisture tests were performed on the scrubber inlet and outlet streams, and on the analyzer sample stream immediately following the knockout trap in the DuPont analyzer (131).

## Conesville Unit No. 5

### Plant and FGD System Description—

The Conesville Generating Station is owned and operated by the Columbus and Southern Ohio Electric Company. Unit 5 is a coal-fired unit rated at a net capacity of 375 MW. The steam generator is a controlled circulation, single reheat, balanced draft boiler with a primary steam flow of 3,130,000 pounds per hour at 2620 psig and 1005°F. The turbine generator is nominally rated at 403 MW. The fuel supply is raw (unwashed) coal as delivered direct from the mine. Typical coal analysis is 14.0 - 16.0 percent ash, 3.5 - 4.5 percent sulfur, with a high heating value of 10,400 to 10,800 Btu/lb.

The flue gas passes from the boiler to a cold-side electrostatic precipitator designed to achieve 99.65 percent particulate removal, induced draft fans, and then to either the flue gas desulfurization system or the full-flow bypass before final exhaust through an 800-foot acid-resistant, brick-lined stack.

The FGD system consists of two mobile-bed scrubber modules. Each module is designed to handle 60 percent of the total full-load flue gas flow (total flow is 1,202,000 cfm at 286°F) with a minimum $SO_2$ removal efficiency of 89.6 percent. The flue gases are contacted countercurrently in a single stage spherical-packing mobile bed with thiosorbic lime slurry. The scrubber effluent slurry is recycled with a bleed-off to a thickener. The thickener concentrates the solids from 9 percent to approximately 30 percent solids. The thickener underflow is pumped to a treatment plant for sludge fixation and disposal. The thickener overflow is recycled to the scrubber presaturator and lime slaker. The Unit 5 scrubbers were placed in service in February 1977 (133).

### Continuous Emissions Monitoring Test Results—

Continuous emissions monitoring was conducted at Conesville Unit No. 5 from June through December 1978. During the total of 184 days that the CEM program was in place, only 34 full data days resulted (boiler and both modules in operation, and at least 18 hours of valid data). The average 24-hour $SO_2$ removal efficiency for these days was 89.2 percent with a standard deviation of 3.6 percent (133).

Intermittent operation at Conesville was due to both the boiler and the scrubber modules. Of the 184 days, the boiler was down for 22, Module A for 102, and Module B for 99. Days with less than 18 hours of valid monitoring accounted for further data attrition.

### Continuing Emission Monitoring Systems—

A DuPont Model 460 $SO_2$ analyzer and a Thermox WDG III $O_2$ analyzer were used at Conesville Unit No. 5. The sample for $O_2$ analysis was withdrawn from the $SO_2$ analyzer so that $SO_2$ and $O_2$ were measured in the same gas stream. The analyzers operated on a 10-minute cycle. $SO_2$ data were recorded by a two-pen recorder, and $O_2$ data were recorded by a single-pen recorder. Moisture data were collected manually approximately every week (133).

## Bruce Mansfield Unit No. 1

### Plant and FGD System Description—

The Bruce Mansfield plant is a three-unit, 2751-MW (gross), coal-fired power generating station. It is owned by the Central Area Power Coordination Group (CAPCO), made up of the Ohio Edison Company, Pennsylvania Power Company, Cleveland Electric Illuminating Company, Duquesne Light Company, and Toledo Edison Company. Bruce Mansfield Unit 1 was placed in service on December 11, 1975, and was placed in full commercial operation on June 1, 1976.

Bruce Mansfield Unit 1 is rated at 917 MW (gross) and fires a high-sulfur, eastern, bituminous coal having an average sulfur content of 3.0 percent and an average ash content of 12.5 percent. The average heating value is 11,500 Btu/lb. To meet emission regulations promulgated by the Commonwealth of Pennsylvania, each unit, including Unit 1, is fitted with a wet lime scrubbing system for the control of particulate and sulfur dioxide.

Flue gas exits from the two air heaters of each boiler at 3.4 million acfm and approximately 285°F, then enters a manifold that distributes it to six separate inlet ducts. Each duct serves one scrubbing train consisting of a variable-throat venturi scrubber module, a 6700-KW (9000 hp) induced-draft fan, and a fixed-throat venturi absorber module. Six scrubbing trains are required on each unit for full boiler load operation. They are arranged in two groups of three, and the treated gas discharged from the three trains in each group flows together into an oil-fired reheat chamber. The heated exhaust gas is then discharged to the atmosphere through a 950 foot stack. The stack contains four separate carbon steel flues, each of which receives the reheated gas stream discharged from one reheat chamber.

The flue gas first enters the top of the venturi scrubber, then passes down and around the adjustable plumb bob and accelerates to a velocity of approximately 200 ft/s through the throat area. The gas is contacted in a cocurrent fashion with lime slurry that is recycled from the base (internal recycle tank) of the scrubber. A tangential-feed arrangement of the slurry beed nozzles atop each scrubber provides the primary sprays that wet the plumb bob and throat area. The incoming gas impacts upon this curtain of slurry spray, forming fine droplets which intimately mix with the gas stream as they pass through the throat area. The gas-slurry stream separates in the lower section of the scrubber. The gas turns 180 degrees and passes upward through mist eliminator. The spent slurry droplets from the sprays are collected in the internal recycle tank for recirculation through the scrubbing circuit. The cleaned gases pass through the single-stage mist eliminator and then flow through the discharge duct to the induced-draft fan.

The induced-draft fan, which overcomes draft losses in the boiler and scrubbing system, receives the saturated gases from the scrubber. The gases then enter the top of the venturi absorber and pass down through the fixed-throat area. In a manner similar to that described for the venturi scrubber, the gas accelerates to a velocity of 100 ft/s through the throat area, where it is contacted with lime slurry in a cocurrent fashion. The slurry, which is recycled from the absorber's internal recycle tank, is sprayed into the gas stream through a tangential-feed arrangement of the slurry feed nozzles. The nozzles spray slurry onto the convering throat area, gas baffles, and center cone of the absorber. The gas-slurry stream separates in the lower section of the absorber. The gas turns 180 degrees and passes upward through the mist eliminator. The spent slurry droplets from the primary and secondary sprays are collected in the internal recycle tank for recirculation through the scrubbing circuit. The cleaned, saturated gases then pass through another mist elimination stage, from which they are sent to the reheat chamber for temperature elevation before discharge to the atmosphere through the stack.

The spent slurry from the venturi absorber is discharged as a continuous bleed stream off the recycle line to the venturi scrubber. The absorber bleed stream enters directly into the internal recycle tank of the scrubber, where it is combined with the scrubber recycle stream and then recirculated through the scrubbing circuit. The spent slurry from the venturi scrubber is also discharged as a continuous bleed stream off the recycle line to a thickener. The spent slurry, which is 8 to 10 percent solids, combines with fly ash slurry from the boiler and then flows into a 200 foot diameter thickener.

The underflow from the thickener, which is 25 to 35 percent solids is then pumped to the waste disposal system for treatment and ultimate disposal. The thickener overflow is used to maintain liquid levels in the scrubber and absorber vessels (134).

Continuous Emissions Monitoring Test Results—

Continuous emissions monitoring was conducted at Bruce Mansfield between September and December 1977. A twenty day run produced an average $SO_2$ removal efficiency of 80.6 percent with a standard deviation of 4.6 percent. A second run of eleven days achieved slightly better results, an average $SO_2$ removal of 85.0 percent with a standard deviation of 3.4 percent (131).

Major areas of concern affecting the Unit 1 FGD system efficiency and reliability were the performance of the scrubber mist eliminators, excessive water entrainment and carryover out of the chimney, pH measurement and control, water balance, reheat burner performance, excessive maintenance associated with the wet induced-draft fan housings, and chimney flue liner failures.

Continuous Emission Monitoring Systems—

At the FGD inlet manifold, $SO_2$ was monitored with a DuPont Model 460 two-point analyzer. The instrument operated on a 10-minute cycle, where each of two measurement points was sampled for five minutes.

Two DuPont Model 460 four-point dual-range anlayzers monitored $SO_2$ at the FGD system outlet. Each instrument measured three module outlets (for the total of six) and the common reheater outlet. Each instrument operated on a 12-minute cycle. $SO_2$ measurements were printed on chart recorders located in the scrubber control room.

For each of the three $SO_2$ analyzers there was a Taylor OA137 paramagnetic $O_2$ analyzer with a chiller-condenser system for sample-gas drying. The sample for $O_2$ analysis was withdrawn from the DuPont instrument downstream of the measurement cell. The moisture contents of the sample streams were determined manually (131).

## Shawnee (TCA)

### Plant and FGD System Description—

The Shawnee Power Station was described previously under the limestone system section. The turbulent contact absorber (TCA) is a tower containing one, two or three beds of low-density spheres, nominally 1 - 1/2 inches in diameter, that are free to move between retaining grids. The incoming flue gas contacts the slurry in these beds, and both $SO_2$ and particulates are removed. The gas passes through a mist eliminator before exiting the TCA. It is reheated prior to discharge through the stack (123).

### Continuous Emissions Monitoring Test Results—

In 42 complete days of continuous emissions monitoring, the Shawnee TCA system using a lime slurry achievd 88.6 percent $SO_2$ removal with a standard deviation of 2.2 percent. Other testing using both lime and limestone slurries enhanced with adipic acid, showed the TCA capable of achieving greater than 95 percent removal of $SO_2$ (133).

### Continuous Emission Monitoring Systems—

The instrumentation at Shawnee was described earlier in the limestone systems descriptions.

## Pleasants Unit No. 1

### Plant and FGD System Description—

Allegheny Power Systems Pleasants Power Plant Unit No. 1 is rated at 640 MW. It fires eastern bituminous coal containing about 3 percent sulfur. Pollution controls consist of a cold-side, high efficiency electrostatic precipitator (ESP) that removes more than 99 percent of the particulate matter, followed by a lime FGD system. After passing through the ESP, the flue gas is collected under positive pressure in a common plenum, which feeds the FGD system's four modules (modules A through D). Two bypass-reheat ducts also draw from the plenum.

The flue gas is introduced into each lime scrubber module through a quencher (a venturi throat with scrubber contactor fluid aspirated into the stream). It then passes through a frothing bed at the base of the (unpacked) spray tower. There are 32 spray nozzles in each module. Each nozzle sprays about 250 gpm of lime slurry absorber fluid into the countercurrent flue gas stream.

Immediately after the spray nozzles, the flue gas stream in each module is further divided between two parallel mist eliminators (eight in total). Shortly thereafter, the bypass reheat duct enters, and the recombined flue gas stream is ducted into a 1000-foot stack (135).

### Continuous Emissions Monitoring Results—

The monitoring program at Pleasants Unit No. 1 began on July 1, 1981 and continued through October 10, 1981. Table 59 below summarizes the $SO_2$ removal efficiency data.

TABLE 59. SUMMARY OF TEST RESULTS AT PLEASANTS UNIT NO. 1

| Module | Number or Data* Days | 24-Hour Average $SO_2$ Removal (%$\mu+\sigma$) |
|---|---|---|
| A | 67 | 94.1 ± 3.6 |
| B | 23 | 92.6 ± 2.6 |
| C | 11 | 95.6 ± 0.7 |
| D | 43 | 89.3 ± 5.5 |

*18 or more valid hours of data within the calendar day.

Module A ran throughout the CEM program. Modules B, C, and D were alternately brought on-line and taken off-line mainly because of maintenance requirements. The scrubber system was designed to have only 3 of the 4 modules in service at any one time. The boiler load during the CEM period varied between 338 and 595 MW.

When operating properly, all scrubbers modules consistently removed more than 90 percent of the $SO_2$. It was observed that the removal was higher at higher gas flows, even at the maximum design flow. This performance was presumably due to the flow dependent behavior of the venturi and frothing bed. Each device was designed to operate most efficiently in a certain flow range, and under low gas flow conditions, either improper aspiration or frothing may have resulted. The only major incident of sustained sub-par scrubber performance occurred in late August; during this period, scrubbing efficiency percentage on Module D was in the eighties because a frothing bed tray had become dislodged.

Minor (and temporary) problems with the scrubbers were noted involving pluggage of sprayer nozzles and pH excursions of the contactor fluids. The inconsistent flows through the nozzles resulted in a temporarily fluctuating stratification pattern of the flue gas in the ducts immediately downstream of the scrubbers. Overall modular performance decreased no more than a few (<10) percent and typically lasted from several days to a few weeks. pH excursions were usually due to either electronic sensor malfunction or mechanical addition problems. The conditions were usually corrected in a day or two and overall FGD system performance was reduced to no more than a few (<5) percent.

The initial goal of the program described here was to monitor long-term performance of the scrubbers at Pleasants No. 1. However, the monitoring system had been previously installed so that it could monitor only-half of each module. Thus, the data resulting from this study actually describe the performance of only one-half of each scrubber module. During the middle part of the monitoring program, EPA Method 6 (time integrated $SO_2$ measurements) runs were made simultaneously on the two halves of either Module A or D in an effort to compare their relative performance. The conclusion from these studies was that the CEM half usually exhibited somewhat lower $SO_2$ scrubbing efficiency than the non-CEM half (135).

## DUAL ALKALI SYSTEMS

Two dual alkali systems have been tested by the EPA in support of the NSPS. These units are A.B. Brown 1 and Cane Run 6. Descriptions of each unit and results are presented below.

### A.B. Brown Unit No. 1

#### Plant and FGD System Description--

A.B. Brown Unit No. 1 is operated by Southern Indiana Gas & Electric Company. It is a pulverized coal-fired unit with gross capacity of 265 MW. The boiler was put into commercial service in 1979.

A.B. Brown Unit No. 1 fires bituminous coal with a heating value of approximately 11,500 Btu/16. The nominal coal sulfur content is 3.5 percent. However, during the continuous emissions monitoring test program described herein, the coal sulfur content averaged 3.22 percent and varied from a low of 1.88 percent to a high of 4.13 percent.

Design flue gas flow from the boiler is 910,000 acfm at 310°F and four percent $O_2$. Dual register burners are mounted on opposite walls of the furnace for limiting $NO_x$ formation.

Particulate matter is removed from the flue gas by a cold-side electrostatic precipitator. $SO_2$ is subsequently controlled by a concentrated mode dual alkali scrubber system, which consists of two three-stage modules. The system was designed to meet the 1971 New Source Performance Standards (1.2 lbs $SO_2$/MMBtu) when burning 4.5 percent sulfur coal by treating all the flue gas at an efficiency of about 85 percent. A.B. Brown Unit No. 1 normally fires a

3.5 percent sulfur coal and has reportedly been able to meet the NSPS by treating 90 percent of the flue gas at 90 percent efficiency while bypassing the remaining ten percent.

$SO_2$ is absorbed in two parallel disc contactor scrubber towers (north and south modules). Inputs to the FGD system consist of water, sodium carbonate, and lime. A single sulfite regeneration system works in concert with the two absorbers. It consists of a lime reactor, thickener, filters, and a surge tank. Inputs to the FGD system consist of water, sodium carbonate and lime.

In the sulfite regeneration system, lime is reacted in a low residence time (~10 minutes) stirred tank. Next, the thickener is gravity fed from the lime reactor. Regenerated liquor then overflows from the thickener to a surge tank, and a centrifugal pump transfers it to the two absorbers (136).

Continuous Emissions Monitoring Test Results—

Continuous emissions monitoring took place at A.B. Brown Unit No. 1 between July 31 and October 8, 1981. Modular FGD efficiency statistics are summarized below in Table 60.

TABLE 60. SUMMARY OF TEST RESULTS AT A.B. BROWN UNIT NO. 1

| Module | Hours of Data | $SO_2$ Removal Efficiency | | |
|---|---|---|---|---|
| | | $\%\mu\pm\sigma$ | Min | Max |
| North | 1,367 | 88.44 ± 3.48 | 75.8 | 97.9 |
| South | 1,250 | 87.03 ± 2.99 | 75.8 | 94.3 |

In terms of 24-hour averages, the north module achieved 54 calendar days with 18 or more valid hourly observations. The average efficiency was 88.2 percent with a standard deviation of 4.3 percent. The south module similarly achieved 48 days of data with resulting average efficiency of 86.1 ± 4.7 percent. The average boiler load during CEM was 129 MW (136).

Continuous Emission Monitoring Systems—

Continuous monitoring data were acquired using the EPA mobile laboratory and the following measurement sensors:

- Horiba PIR-2000 (NDIR) for monitoring $SO_2$ at the FGD system inlet,
- DuPont Model 400 (NDUV) for monitoring $SO_2$ at the FGD system outlet, and
- MSA 802 (paramagnetic) for monitoring $O_2$ at both inlet and outlet.

The CEM system monitored one FGD module at a time, cycling between the two every 15 minutes by properly placing the respective valves contained in a valve switching box in the proper position. The instruments were stringently calibrated on a weekly basis during the monitoring period.

Monitoring data were recorded continuously by a computer system and stored on a floppy disk. An Autodata 980 logger served as a backup to assure a high rate of data capture. Final data reduction was conducted using a time-sharing computer (136).

## Cane Run Unit No. 6

### Plant and FGD System Description—

Louisville Gas and Electric Company's Cane Run No. 6 is a 300 MW gross peak capacity tangentially-fired, dry bottom boiler built by Combustion Engineering. It was placed in service in 1969. It is nominally rated at 280 MW and has a continuous rating of 300 MW. Its minimum controllable rating is 60 MW, and its annual average load has been about 180 MW.

The furnace of Cane Run No. 6 is tangentially-fired by five elevations of coal burners located near the furnace corners. This arrangement directs the fuel streams to imaginary firing circles in the center of the furnace in order to produce efficient mixing of fuel and air. Pulverized coal and primary air are delivered at about 82°C from the classifiers to the burners.

Cane Run No. 6 normally fires Western Kentucky No. 9, a high slagging Eastern Bituminous coal. This coal has an average heating value of 11,000 Btu/lb, ranging from 10,500 to 12,000 Btu/lb. Moisture content ranges from 8.0 to 10.75 percent with an average of 8.95 percent. Ash content ranges from 15.5 to 24.5 percent with an average of 17.1 percent. And sulfur content ranges from 3.5 to 6.3 percent with an average of 4.8 percent.

Two single-stage electrostatic precipitators in parallel control particulate emissions. Design efficiency of the precipitators is 99.4 percent.

$SO_2$ is controlled by a dual alkali FGD system, which started operation in October 1978. Two absorber trains operate in parallel, fed with flue gas from the two ESPs. The system is closed loop; the only direct process waste is a washed filter cake of calcium-sulfur salts. No other liquid or solid process streams are discharged.

Each train in the scrubbing subsystem consists of a countercurrent tray tower absorber (scrubber) equipped with a liquid entrainment separator, an enclosed recycle tank, and recycle pumps. Flue gas from the electrostatic precipitator is forced by booster fan through the scrubber. The gas passes upward through sprays, two trays, and finally chevron demisters. The scrubbed flue gas leaving the tower is mixed with hot gas from an oil-fired reheater before entering the stack.

Liquor fed to the top tray of the scrubber passes across the trays and down the tower, and collects at the bottom of the unit where it is mixed with sodium carbonate makeup solution. The spent liquor, having absorbed sulfur dioxide from the gas, contains sodium sulfite and bisulfite. Some of this spent liquor is recycled to the sprays and some to the top tray for pH control. The rest is sent to the absorbent regeneration subsystem.

The spent scrubber solution, which is continuously bled from the scrubber recycle stream, is sent to the absorbent regeneration subsystem. Here, the solution is reacted with hydrated lime in a series of two lime reaction tanks. The lime reduces the acidity of the bisulfite in the scrubber bleed, and converts sodium sulfite and sodium sulfate to sodium hydroxide. These reactions precipitate a mixture of calcium sulfate and calcium sulfite solids.

Slurry from the lime reaction tanks is fed to the center well of the thickener. Thickener underflow is dewatered on rotary drum vacuum filters. Filter cake is washed by a series of water sprays which remove most of the occulded soluble sodium salts from the cake. This filtrate is sent back (via the thickener) the the scrubbing subsystem, conserving sodium values. Solids dewatering produces the only process waste stream from the scrubbing system, the washed filter cake. A small volume liquid discharge is also produced in this subsystem. This discharge consists of condensate collected from vacuum pump silencers (137,138).

### Continuous Emissions Monitoring Test Results—

CEM testing of the Cane Run Unit No. 6 dual alkali FGD system was conducted over a thirteen month period from May 1980 through May 1981. Average $SO_2$ removal over this period was 91.6 percent (based on 12 monthly excluding December 1980, averages). The FGD system operated 7,005 hours and exhibited excellent reliability (hours operated/hours required x 100) of 94.1 percent. The boiler did not usually operate at maximum capacity during the thirteen months of testing.

The lowest monthly $SO_2$ removal efficiency, 82.6 percent, occurred in November 1980. This poor performance was largely attributed to the presence of significant amounts of suspended solids in the thickener overflow. Solids carryover from the thickener overflow increased the rate of absorber plugging and reduced the $SO_2$ removal capability.

In December 1980, the FGD system operated only for 14.4 hours of the 240.4 hours that the No. 6 boiler was in operation. The boiler was shut down during much of that month, primarily for scheduled maintenance. During this outage the FGD system was considered available. However, after the boiler came back on-line, a thickener rake problem rendered the FGD system inoperable for 297.6 hours. The rake in the thickener had separated from the drive shaft. The FGD system and the thickener were briefly operated (14.4 hours) to keep the thickener slurry from freezing and were then shut down so the rake could be repaired. No system characterization tests were performed that month.

Other recurrent problems included piping failure due to vibration or mechanical stress, gas leakage through the scrubber bypass damper, moisture entering the soda ash silo, and a high incidence of torn vacuum filter cloth (139).

### Continuous Emission Monitoring Systems—

A DuPont Model 460 was used to monitor $SO_2$ at the inlet. A Lear Siegler SM-810 was used at the outlet. To monitor $O_2$, a Hagen/Westinghouse electricatalytic analyzer was used at the inlet, and a Lear Siegler CM-50 at the outlet. Acquisition and reduction of data was done automatically (139).

## WELLMAN-LORD SYSTEM

Only one Wellman-Lord system, the unit at D.H. Mitchell 11, has been tested with continuous monitors. The results are discussed below.

## D.H. Mitchell Unit No. 11

### Plant and FGD System Description—

D.H. Mitchell Unit No. 11 is owned and operated by Northern Indiana Public Service Comapny (NIPSCO). Four generating units are located at this station. Coal is the primary fuel for each.

The Unit No. 11 boiler is a balanced draft, dry-bottom, wall-fired radiant boiler rated at 115 MW. Normal output ranges from 46 to 115 MW with an average load factor of approximately 80 percent or 92 MW.

High volatile bituminous coal normally fires Unit No. 11. Typical coal analyses are 11,000 Btu/lb, 13.6 percent moisture, 9.5 percent ash, and 3.2 percent sulfur.

In 1976, Unit No. 11 was retrofitted with a Wellman-Lord FGD system integrated with an Allied $SO_2$ reduction process. The FGD unit was designed to remove 90 percent of the $SO_2$ delivered with the flue gas at flue gas rates equivalent to a boiler load of 92 MW (80 percent load factor) while burning coal with a nominal sulfur content slightly above three percent. The major process steps for the flue gas are booster fan, prescrubber, absorber, and stack. $SO_2$ removed in the absorber is recovered and sent to the reduction unit.

The prescrubber is a single-stage orifice contactor for removing additional particulate matter. A pump recirculates the scrubber water from a sump back to the contactor. In order to control a solids buildup in the liquid stream, a purge stream is withdrawn; makeup water is added to the prescrubber to compensate for this loss and to humidify the flue gas. This purge stream is sent to the power station's fly ash settling ponds.

The cooled, humidified flue gas leaves the prescrubber and enters the bottom of a three stage absorber where the gas is contacted with the sulfite/bisulfite solution flowing contercurrently to the gas stream. The solution absorbs the $SO_2$ and the treated flue gas is then discharged to the atmosphere through a stack.

The spent sulfite/bisulfite solution is removed from the bottom tray of the absorber and sent to a surge tank for storage prior to regeneration in the $SO_2$ recovery step. During recovery of the $SO_2$, the spent absorbent is regenerated in a steam-heated, single-effect evaporator and is then returned to the absorber feed tank. The surge tank and absorber feed tank provide surge capacity for operating for limited time periods at flue n.as rates in excess of 92 MW equivalent. To prevent accumulation of sodium sulfate in the absorbing solution stream, a purge stream is sent to the purge treatment area. Here, the purge stream is crystallized and centrifuged and the solid product is removed and dried, yielding a salable sulfate by-product. The sodium values lost in the purge stream are made up by adding $Na_2CO_3$ to the regenerated sulfite/bisulfite solution.

$SO_2$ released in the evaporator is taken overhead and sent to the $SO_2$ reduction area. The reduction step is a proprietary process that utilizes natural gas ($CH_4$) for the reduction of $SO_2$ to $H_2S$ and, ultimately, to elemental sulfur in molten form. A small stream of tail gas is returned after incineration to the inlet of the booster blower (84,140).

Continuous Emissions Monitoring Test Results--

The performance of the Wellman-Lord FGD system at D.H. Mitchell Unit No. 11 has been monitored over several years time. Two years of EPA sponsored demonstration testing began in September 1977.

During the first demonstration year, the full FGD plant operated a total of 2,155 hours (140). Partial operation, with the reduction unit down and minimal recovery of $SO_2$ or with the bypass damper open, occurred a total of 1,681 hours, for a total operating time for the absorber/evaporator of 3,836 hours. $SO_2$ removal efficiency averaged 89 percent during those hours at an average inlet $SO_2$ concentration of 2,081 ppm. The boiler load averaged 75 MW (gross) during the period. The FGD system reliability (hours of full operation/hours required x 100) was 50 percent.

Problems that were encountered from the start of the first demonstration year prevented consistent operation of the FGD plant until the last two months of the year. The problems were primarily boiler related or problems at the boiler/FGD plant interface. The major problems were:

- Coal feeding and coal quality — Inconsistent coal quality caused flucuating coal feed rates. This led to unstable flue gas rates and steam pressures, conditions under which the FGD plant was unable to operate.
- Booster blower — The primary problems were rapid deterioration of the fan blades from contact with the wet flue gas and flyash, imbalance of the fan due to flyash buildup, and problems with blower and turbine controls and lubrication system.
- Isolation damper — A guillotine damper, installed in the flue gas duct upstream of the booster blower, isolates the FGD plant from the boiler. Fly ash hardened in the damper tracks and prevented opening or closing when needed.
- Absorber leaks — Leaks occurred at the bottom collector tray of the absorber.

The second year of demonstration testing began in October 1978 and continued through October 1979 (140). The FGD unit achieved 5,079 hours of full operation plus another 1,353 hours of partial operation during that time. The average $SO_2$ removal efficiency was 89.6 percent with a standard deviation of less than one percent. Boiler load generally ranged between 50 and 85 MW. The average inlet $SO_2$ concentration was 2,097 ppm. System reliability (based on hours of full operation) was 61 percent.

Several pieces of equipment were responsible for most of the FGD systems' problems. The reduction unit caused 59 days of interruptions; the most frequent outages were those due to sulfur deposition, leaks, and valve repairs. Problems with the evaporation circulating pump caused 28 days of interruption. This pump circulates spent absorber solution through the evaporator heater. Problems with other equipment, including the absorber, booster blower, and evaporator casued a total of 36 days of interruption.

### Continuous Emission Monitoring Systems—

Analysis of flue gas $SO_2$ was accomplished using the DuPont Model 400 UV spectrophotometer. $O_2$ was monitored using the Beckman Model 742 analyzer. The data acquisition system (DAS) consisted primarily of a control unit, a minicomputer, a nine-track tape system, and a printer. Six strip chart recorders were available for backup in case of a DAS failure. The DAS had the capability of storing data on magnetic tape. However, hardware difficulties with the tape transport unit were experienced such that very little automated data reduction was possible. The hard-copy data from the printer was relied on for most data processing and analysis (133).

## MAGNESIUM OXIDE SYSTEM

The Eddystone prototype magnesium oxide system was tested by the EPA before construction was started on the full scale units. This is currently the only planned installation of this technology.

### Eddystone Unit No. 1

### Plant and FGD System Description—

Unit No. 1 at Philadelphia Electric Company's Eddystone station is rated at 325 MW generating capacity. It burns West Virginia-Pennsylvania bituminous coal with the following average values: heating value, 13,600 Btu/lb; moisture content, 5.9 percent; ash content 9.4 percent; and sulfur content, 2.6 percent.

Flue gas from the boiler passes through a mechanical collector and an electrostatic precipitator (ESP) for particulate control. Following the ESP the total flue gas flow of 927,000 acfm is split into three streams, each equpped with a venturi scrubber where the flue gas is contacted with river water for additional particulate removal. In this step the gas is cooled from 300-330°F to 125-130°F.

One of the three venturi scrubbers is followed by an $SO_2$ removal absorber. This $SO_2$ scrubber is equipped with two scrubbing stages (venturi rod modules) and two stages of mist elimination. Magnesium oxide slurry is used as the $SO_2$ removal absorber.

Stack gas reheat is accomplished by direct firing. Flexibility of operation is incorporated in the ductwork through systems that permit bypassing the $SO_2$ scrubber, each train, or the entire particulate/$SO_2$ scrubber plant.

In the $SO_2$ scrubber the flue gas emerging from the particulate scrubber is contacted with a 5-10% aqueous slurry of magnesium sulfite ($MgSO_3$) to remove better than 90 percent of the sulfur dioxide in the flue gas. In this step the insoluble magnesium sulfite is converted to soluble magnesium bisulfite ($Mg(HSO_3)_2$). In the scrubber surge tank a slurry of slaked magnesium oxide (MgO) is added to the circulating scrubber liquor at a rate sufficient to maintain the pH at about 6.3. The addition of MgO converts the $Mg(HSO_3)_2$ to insoluble $MgSO_3$. A bleed stream from the circulating scrubber slurry goes to a thickener, centrifuge, and rotary kiln circuit, where the $MgSO_3$ is recovered as a dry product. The mother liquor recovered from this operation is used to slake the MgO and to wash the scrubber mist eliminators.

The dry $MgSO_3$ is calcined in a fluid bed reactor in the off-site regeneration facility. This facility includes an air preheater, cyclone separator system, venturi scrubbing, gas cooling tower, drying tower, and mist eliminator for separation of the MgO crystals, and the cleaning, cooling and drying of the $SO_2$-rich gas for transfer to the sulfuric acid plant (89).

### Continuous Emissions Monitoring Test Results—

A limited amount of CEM testing sponsored by EPA was conducted at Eddystone Unit No. 1. Testing was discontinued in late December 1977. The results of eight 24-hour periods on one of the two FGD modules indicated an average $SO_2$ removal efficiency of 96.8 percent with a standard deviation of 1.2 percent. The boiler was operating at near capacity during testing (127).

Operating problems with the FGD system included MgO slaking, plugging and leaking of rotary screw feeders, and leaking of slurry circulating pumps.

### Continuous Emission Monitoring Systems—

Continuous emissions monitoring at Eddystone was handled by a single DuPont Model 460 dual-range five-point $SO_2$ analyzer. This instrument measured $SO_2$ at the absorber inlet and outlet. It normally operated on a 20-minute cycle. Measurements were printed on a multipoint chart recorder located in the scrubber control room.

Continuous $O_2$ measurement was not conducted at Eddystone. Manual procedures were used instead. The moisture content of the sample streams was determined by manual measurements at each sample site (131).

## SPRAY DRYING SYSTEM

No utility spray drying systems have been tested as yet by the EPA. This is because of the very limited number of systems which have started up. It is most likely that these systems will be tested in the near future as commercial systems come on-line. The results of an industrial boiler spray drying system test are presented below.

## Celanese Industrial

### Plant and FGD System Description—

The Celanese Amcelle synthetic fibers plant at Cumberland, Maryland produces acetate fibers that are used in a wide variety of consumer goods. Steam requirements at the plant include process steam, space heating, and hot water. Celanese replaced an existing gas/fuel oil-fired boiler with a coal-fired one in 1979 to achieve fuel cost savings.

The boiler is of the spreader-stoker type and burns eastern coal having an average heating value of 12,400 Btu/lb, ash content of 14.9 percent, and sulfur content of 1.9 percent. The boiler has a design capacity of 110,000 lb/hr of steam at 700°F and 725 psig.

The gas cleaning system consists of a single 20 foot diameter spray dryer in series with a four-compartment, pulse-type fabric filter (bag house). The system includes lime unloading, storage, and conveying equipment, lime slaking and feed preparation equipment; connecting ducting, and associated auxiliary equipment and controls.

The spray dryer centrifugally disperses a lime slurry into the flue gas. Sulfur dioxide is absorbed on the droplets and reacts with the calcium hydroxide to form calcium sulfate. The atomized droplets rapidly evaporate in the spray dryer, resulting in a desulfurized, cooled, and humidified solids-bearing gas.

The resulting mixture of scrubbed combustion gases, fly ash, dry reaction products, and unreacted sorbent flows from the spray dryer into the four-compartment fabric filter where the solid material contained in the flue gas stream is collected on the filter bags. Additional $SO_2$ removal occurs on the filter bags because of the presence of unreacted lime in the filter cake. The cleaned flue gas is then discharged through the fabric filter plenum to the induced draft fan and stack.

Approximately 20 percent of the flue gas exiting the boiler bypasses the spray dryer and is combined with the treated gas leaving the spray dryer. This provides the reheat necessary to raise the temperature of the gas stream to a safe margin above the adiabatic saturation temperature at the spray dryer outlet, thus minimizing condensation problems in the baghouse.

The primary controls on the spray dryer limit $SO_2$ concentration at the exit of the baghouse and adjust the spray dryer exit temperature (the latter to control flue gas temperature at the inlet to the baghouse so as to prevent bag damage). Spray dryer exit temperature is automatically controlled by metering the water feed to the spray dryer. The outlet $SO_2$ concentration from the baghouse is controlled by the use of a feedback control system. As $SO_2$ stack gas analyzer provides the input signal and an electronic controller varies the feed rate of the pumps supplying the lime feed to the spray dryer. An override system is provided to allow manual control of the supply pumps when the feedback control is inoperative or malfunctioning (141).

### Continuous Emissions Monitoring Test Results—

The continuous emissions monitoring program began in May 1980. Because of problems with the boiler, the FGD system, and the CEM system, the collection of CEM data was confined to the period between August 28, 1980 and September 30, 1980.

A total of 23 days of CEM data was obtained with at least 18 individual data hours. The average $SO_2$ removal efficiency of the spray dryer/baghouse system was 70 percent with a standard deviation of 9 percent. Over the last six days of the program, after several operational difficulties had been resolved, $SO_2$ removal efficiency averaged 78.5 percent (141).

The boiler operated at about half its rated load with relatively steady steam production for most of the CEM period.

Several FGD system operational problems were noted during the program. One was caused by the variability of coal quality. Coal sulfur content varied widely throughout the early part of the program. The quality became less variable near the end of the program, but proximate analyses of daily coal deliveries gave sulfur contents ranging from 1.25 to 2.76 percent. When operating in an automatic control mode to keep the outlet $SO_2$ concentration at a set value, the FGD system responded to rapid changes in inlet $SO_2$ concentration within an hour so that hourly averages of emissions remained constant. Under manual control, the outlet $SO_2$ concentration followed the inlet concentration in the absence of operator adjustment. With uniform coal quality and automatic control of slurry flow, large fluctuations in the inlet $SO_2$ concentration were absent and a steady outlet $SO_2$ concentration was maintained.

Another coal problem involved the amount of fines in the coal. Coal fines, when suddenly dumped into the furnace, caused rapid changes in boiler load and flue gas flow, changes in $SO_2$ emissions, as well as increased particulate matter and opacity levels in the stack. Fast changes in flue gas flow and $SO_2$ concentration made it difficult for the spray dryer to keep $SO_2$ emissions at a desired level.

The lime feed system also caused FGD system difficulties. It uses a paste-type slaker with a grit-screen removal system and is designed to supply sufficient slurry for $SO_2$ removal from flue gas when burning coal with up to 2 percent sulfur. The slurry is fed to the atomizer by progressive cavity pumps. Under the design conditions, one pump is in use and one is a spare. However, to cope with the higher sulfur coals encountered, the single pump had to be operated at high speeds and this led to rapid pump wear. To alleviate this, the system was modified to use both pumps in parallel operation. This was the normal mode of operation throughout the latter part of the test period.

Some other problems with the lime feed system related to plugging somewhere in the system because of grit in the lime. Although grit was supposed to have been removed by a screen inside the slaker, damaging quantities passed through or bypassed the screen into the rest of the system. Failure to remove grit caused plugging inside the flow lines, in the slurry pump, and in valves at various times, in addition to excessive wear in the pumps. Dual element screen filters were eventually installed in the feed system, but not enough time elapsed before the end of the program to assess whether they solved the problem. It was reported later that this solution indeed resolved the problem.

There were three types of problems that occurred within the spray dryer. One of them resulted in the wetting of the dryer wall and discharge of damp material from the dryer. This was caused by maldistribution of lime slurry in the atomizer and was corrected by redesign. The other two problems encountered with the spray dryer also related to the atomizer. The spray machine was subject to clogging with grit particles not screened sufficiently from the slurry. The spray dryer was shut down for cleaning when this clogging occurred. The other problem was early failure of the bearings supporting the shaft of the atomizer wheel, caused by an unbalance of the atomizer wheel.

Other difficulties that affected performance of the FGD system were experienced in the fabric filter (baghouse). The most serious problem was the unexpectedly high pressure drop through the filter. This was apparently caused by moisture on the bags which occurred during an upset and combined with ash and lime to form a coating that increased the resistance to flow. These bags were replaced and modifications were made to increase the pulse-jet air volume by approximately 15 percent to improve particle removal from the bags. Tests since this modification indicate that this has solved the flow-resistance problem. Finally, there was a reduction in $SO_2$ removal efficiency immediately after bag replacement, presumably due to an absence of filter cake containing lime on the new bags.

Because of some of the problems discussed above, redundancy of critical components was recommended in the reference cited in this section. Specifically, it was recommended to have a spare slurry pump and a spare atomizer available. It was also recommended that filters be set up to provide grit-free, uniterrupted slurry flow to the spray dryer during periods when one filter element is being cleaned or replaced.

### Continuous Emission Monitoring Systems--

The collected data included $SO_2$ and $O_2$ concentrations at the inlet and outlet of the FGD system. These flue gas streams were sampled and analyzed continuously, but the resulting data were recorded only four times per hour. These values were then averaged to derive the hourly average value for each parameter.

Flue gas was anlayzed using a Thermo-Electron Model 40 Fluorescent $SO_2$ analyzer, and Beckman Model 755 Paramagnetic $O_2$ analyzer. These instruments were certified according to procedures outlined by EPA.

A data acquisition system (DAS) was used to record emisisons data, which were also transmitted to strip chart recorders used as backup. Data from the DAS were processed manually (141).

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